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Rein, David Arno (M.S., Civil Engineering)

IMPROVED VACUUM FILTRATION OF WASTE WATER SLUDGE WITH FILTER MEDIUM PORE SIZE CONTROL

Thesis directed by Assistant Professor Edwin Bennett

A primary objective of waste water treatment is the separation of the solid and liquid fractions of the waste stream. Once these solids are separated, their disposal must be accomplished. Sludge disposal is a major problem confronting the waste treatment field.

There are many alternative means available for processing waste water sludges. These alternatives must be known and investigated to design the optimum system. The various unit operations which can be used in a waste water solids handling and disposal system are presented to provide a background from which to judge the vacuum filtration operation.

Vacuum filtration is a popular means of dewatering sludges to facilitate final solids disposal. This process involves many complex and interrelated parameters. Vacuum filters utilizing synthetic or metallic filter media produce much higher filtration rates than those filters utilizing fine woven cloth media. Previous investigators have attributed the higher filtration rates to a lowering of the filter medium flow resistance; however, another important phenomenon caused by the difference in the size of the pore diameters of these media is the major factor contributing to the increase filtration rates which are obtained.

The synthetic and metallic filter media have relatively large pore openings which allow various fractions of the fine sludge solid particles to pass through with the filtrate. An investigation of the effects of various degrees of solids passage on vacuum filtration rates revealed that these media actually produce a change in the characteristic specific resistance of the filter cake. This change was found to be the major factor contributing to the increased filtration rates.

The laboratory procedure employed permitted the filtrate flow characteristics to be evaluated for incremental depths of filter cake deposited. The results obtained illustrate the effect of various pore diameter filter media, form vacuum, and initial sludge solids concentration on filtration rates. The results indicate a new idea in vacuum filtration, a variable specific resistance with the filter operating as a solids classifier.

This abstract is approved as to form and content.

Signed Edwin R. Bennett

Faculty member in charge of thesis

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A	Cross-sectional area of bed of solids or filter medium
α	Fraction of cycle time taken up by cake formation
B	Ratio of volumes of elutriating water to volume of liquid in the sludge
c	Volume of solids deposited per unit volume of filtrate
γ_s	Unit weight of solids
C_f	Final cake moisture content
C_i	Initial moisture content of sludge
C', C''	Constants
D	Original alkalinity of sludge before elutriation
D_h	Hydraulic diameter of solid particles
D_s	Geometric mean sieve diameter of solid particles
E	Alkalinity of elutriated sludge
ϵ	Porosity of bed of solids or filter cake
f	Friction factor
f'	3f, friction factor
g_c	Newtons conversion factor
h_f	Frictional head loss
i	Initial sludge solids concentration
K'	A constant
L	Depth of bed or filter cake
m	Fraction of time suction acts
μ	Filtrate viscosity
N	Total number of solid particles
n	Coefficient of compressibility, constant

N_{RE}	Reynolds number
ω	Weight of dry cake solids deposited per unit volume of filtrate
ΔP	Pressure drop across filter cake
φ_s	Particle shape factor
Q	Volumetric flow of filtrate
Q_a	Equivalent flow through filter medium and apparatus
Q_c	Equivalent flow through filter cake
Q_T	Total flow through system
R	Hydraulic radius
R_a	Medium and apparatus resistance to flow
R_c	Cake resistance to flow
R_f	Media resistance to flow
R_r	Specific resistance of the cake $\times 10^7$
R_T	Total resistance to flow
r	Specific resistance of filter cake
r'	r/γ_s
r_o	A constant
ρ	Filtrate density
s	Number of elutriation stages
S_p	Surface area of solid particles
t	Time of filtration or cake formation
t_c	Filter cycle time
V	Volume of filtrate
V_p	Volume of solid particle
V_s	Volume of sludge filtered

W	Alkalinity of elutriating water
Y	Filter yield
z	Ratio of wet cake to dry cake, weight

CHAPTER I

INTRODUCTION

In 1898 Lord Iddesleigh, the Chairman of the Royal Commission in England, was examining Dr. H. Maclean Wilson, then Chief Inspector of the West Reeling of Yorkshire River Board, when he asked this pertinent question: "What do you do with sludge that remains behind at a sewage works?" Dr. Wilson replied, "It is really one of the great problems... enormous quantities of sludge are produced..."(1). Today almost 70 years later, sludge handling and disposal is an even greater problem which still remains to be solved. Although advances in the methods and means of sludge handling and disposal have been significant during the past decade, the increase of the quantity of sludge to be dealt with has been more significant. The increasing population, the trend towards urbanization, the more stringent standards controlling final effluent discharges, and the development of modern secondary waste-water treatment processes have all added to the magnitude of the sludge handling and disposal problem.

The basic objectives of sludge handling and disposal systems are dewatering of the sludge, destruction of the organic fraction of the solids, and disposal of the residue. Vacuum filtration is a mechanical method of dewatering sludges. The basic design of the facuum filter was patented in England by William and James Hart in 1872 (2). It's

first application to waste water sludges in the United States was in Milwaukee in 1925 (3). Although substantial progress has been made in the field of sewage sludge filtration since its first application, the design of the filter remains fundamentally the same in principle as that envisioned by William and James Hart. Only recently have university-sponsored studies begun making contributions to this important method of sewage sludge dewatering.

Importance of the Study

The disposal of domestic waste water sludges has long been a problem for both large and small communities. The most common method utilized in waste treatment plants is anaerobic digestion followed by open bed dewatering. The cost of such a system constitutes a significant portion of the total cost of the treatment plant, and yet, does not provide for the maximum reduction nor the ultimate destruction of the remaining waste organic solids. Today, because of the rising value of land, urbanization, and the demands of the public for a better environment anaerobic digestion is becoming less attractive. Mechanical and chemical operations such as vacuum filtration and incineration are becoming economically competitive with digestion and drying beds as a means of sludge disposal.

The average solids content of a domestic waste water is presented in table 1-1 (4).

TABLE 1-1

AVERAGE PER CAPITA SOLIDS IN DOMESTIC WASTE WATER, GRAMS PER CAPITA PER DAY

State of solids	Mineral	Organic	Total
1. Suspended	25	65	90
(a) Settleable	15	39	54
(b) Non-settleable	10	26	36
2. Dissolved	80	80	160
3. Total	105	145	250

(gram per capita = 2.2 lb. per 1000 population)

These figures indicate that for a domestic waste water one can expect approximately 0.55 lb/cap/day total solids production of which 0.198 lb/cap/day would be suspended solids. These figures, taken in conjunction with the present and projected population figures for the United States present a startling indication of the magnitude of the solids handling and disposal problem. For a population of 190,000,000 the total solids production would be 104,500,000 lbs. per day. Morgan and Thomson state that the enigma of sewage treatment is the disposal of ever-accumulating sludge (5). Bloodgood states that the problem of sludge disposal is as great or greater than that of purifying the sewage (6).

Cost data for solids handling and disposal systems are difficult to cite accurately because most of the operational records at the treatment plants are so limited and vague with respect to unit costs. Hampton states that the disposal of solids removed from domestic waste waters presents an increasingly difficult problem to cities of all

sizes due to high land values and high labor costs (7). The capital cost of the solids handling and disposal system for the Minneapolis-Saint Paul Sanitary District Sewage Treatment Plant was estimated at approximately 25 percent of the total capital cost of the primary and secondary treatment plant (8). Many people believe that sewage sludge can be sold as a fertilizer for a net profit; however, when capital, operating, and maintenance costs are totalled, there is always a net cost. Frazee concludes that sludge handling and disposal is an expensive operation by any method now in use at large installations (9).

The cost figures which can be found in the literature for solids handling and disposal systems vary widely. One of the lowest figures available which included all of the costs associated with the solids system, capital, operating, maintenance, and indirect costs, was at the Minneapolis-Saint Paul Treatment plant where they processed their sludge for $9.90/ton of dry solids in 1962 (8). On the other end of the scale was the cost figure for the New Rochelle treatment plant where they processed their sludge for $62.20/ton of dry solids in 1959 (10). Thus, it can be seen that a well designed solids handling and disposal system is extremely important from a monetary point of view.

The methods utilized for designing vacuum filters present many problems (11). In most cases the same basic filter is used for installations even though it would seem that a specific type of filter would be best suited for a certain sludge. Filters can be sized from data showing quantities of sludge, sludge characteristics, filtration rates, cake moisture, and filter operational variables. The basic parameter

now employed to describe the filtration characteristics of a sludge is the specific resistance. Methods of determining an average value for this parameter are known; however, they do not consider the important phenomena of the initial loss of fine solid particles in the filtrate, a fact which has contributed significantly to the success of the modern coarse medium vacuum filter. Few attempts have been made to predict plant scale operations using laboratory data and the theory of filtration. The usual procedure is to use average data from plants using similar sewage treatment methods on the same types of sewage (12).

Scope of Study

There are many alternative unit operations available to the designer of a solids handling and disposal system. He must be familiar with all of the alternatives if the optimum system is to be designed. The characteristics of the unit operations which are employed in solids handling and disposal systems are described in Chapter Two.

The modern vacuum filters using coarse filter media, such as the (Coilfilter), produce much higher yields than those using fine woven cloth media. The theory of vacuum filtration as generally applied with the standard Buchner Funnel Test predicts a portion of this increase on the basis of lowering the filter media flow resistance; however, it does not account for the increase in yields that are usually produced. The coarse filter media actually produce a change in the filter cake resistance to the flow of filtrate which has not been investigated and which is generally ignored in the Buchner Funnel Test when determining an average specific cake resistance.

The theory of vacuum filtration is developed in Chapter Three with particular emphasis being placed on finding the parameters which determine the specific resistance of a filter cake. The parameters in the standard vacuum filtration equation are investigated and discussed in detail.

In Chapter Four the use of a constant value for the specific resistance parameter in the vacuum filtration equation is investigated. The apparatus devised for the laboratory enables the investigator to determine the change in the specific resistance as the filter cake forms. The variables which are investigated as to their effect on the specific resistance value are filter media, vacuum level, and feed solids concentration. The results of these investigations are discussed and the appropriate conclusions are drawn in Chapter Five.

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CHAPTER II

WASTE WATER SOLIDS HANDLING AND DISPOSAL SYSTEMS

Waste Water Solids

Domestic waste water solids originate from four sources, primary sedimentation, chemical coagulation, activated sludge reactors, and trickling filters. These solids are further classified as raw or digested. Digested sludges are sludges which have been subjected to anaerobic decomposition.

The sludge from primary sedimentation consists of nearly all of the settleable solids from the waste water. It is usually gray in color and gelatinous in consistency. The solids content of a primary sludge will vary from 2-5 percent (4), and will contain a wide range of particle sizes. Primary sludge is best suited for dewatering by mechanical means, although the final moisture content of the dewatered sludge is usually relatively high.

The sludge resulting from chemical coagulation is generally black in color and gelatinous in consistency. Its solids content varies from 2-5 percent depending on the amount of chemicals used (4). The ease with which it can be dewatered by mechanical methods is also dependent on the quantity of chemicals employed. The dewaterability and the concentration of the sludge will increase with increasing chemical concentration up to a certain limit and thence decrease.

The sludge resulting from activated sludge units has a brown flocculent appearance when in good condition and a dark appearance when approaching a septic condition. It is produced by biological reaction in the activated sludge unit. Its solid content will vary from 1-5 percent (4) and will contain uniformly sized particles. It is generally the most difficult type of sludge to dewater by mechanical methods.

The sludge resulting from trickling filters is brown in color and somewhat flocculent in nature. It is produced by biological reactions in the trickling filter. Trickling filter sludge is generally recycled to the primary sedimentation tank and, since it is usually a small quantity, it loses its identity in the primary sludge.

The actual quantity of sludge to be expected from a domestic waste water must be calculated from the composition and quantity of the waste water, the removal efficiencies of the various units, and the sludge production rates of the activated sludge reactor and the trickling filter.

Pretreatment

Sludge thickening is a basic step in the preparation of domestic waste water sludges for final disposal. The objective of the operation is to reduce the volume of sludge by removing water from it. Whatever the final method of handling of the sludge, a reduction in its volume is always advantageous. Sludge thickening can be accomplished by flotation, sedimentation with stirring, or recirculation.

Flotation is generally used as a means of thickening activated sludges (13). Sedimentation with stirring is used as a means of thickening raw and digested sludges. Controlled recirculation of digested sludge into thickened raw primary and modified aeration sludges is a relatively new technique of thickening sludges which is practiced at the Bowery Bay Pollution Control Plant (14). The technique was found to increase the relative storage capacity of the secondary digestors and storage tanks, and to reduce the volume of sludge requiring final disposal. Sludge thickeners are also used to advantage to provide a more uniform sludge prior to mechanical dewatering.

Dewatering

Drying beds

Sand drying beds continue to serve a useful purpose in sludge disposal; however, due to large land area requirements, high labor costs, and the development of more efficient mechanical and heat-drying equipment, they are being limited more to small and medium sized towns (15). Drying beds are confined to the dewatering of well-digested sludges, as the other types of sludges do not drain well and have a tendency to produce odors (4). The beds are made up of graded sand with open-joint tile as underdrains. It is estimated that approximately 20-25 pounds of dry solids per square foot per year can be loaded onto a properly designed sand base drying bed (16). A common design figure for the northern United States is a requirement of one square foot per capita (4).

Centrifugation

Centrifugation is a mechanical means of dewatering sludges which has recently gained acceptance in the sanitary engineering field. Sludge concentration is accomplished by subjecting the underflow to a force of 3000 to 5700 gravities in a solid bowl centrifuge. The factors which made centrifugation ineffective in early treatment plants were its low efficiency and poor solids separation capabilities (17). Modern plants using this method of dewatering are obtaining solids concentrations in the centrifuge cake of 33-50 percent and solids recoveries of 62-92 percent (18). The major advantage claimed for centrifugation is that it can be employed without chemical conditioning of the sludge.

Vacuum filtration

In the United States vacuum filtration is the most popular mechanical means of dewatering waste water sludges. Its rapid gain in popularity during the past few years can be attributed to the relatively new metallic filter media, the use of organic polymers to lower conditioning costs, the location of waste water treatment plants in congested urban areas where drying beds and digesters are undesirable and costly, and the increasing interest in fresh solids dewatering. Sludge dewatering by vacuum filters is usually accomplished by applying a vacuum to a porous drum which is partially submerged in a vat of the sludge to be dewatered. In most operations it is necessary to condition the sludge prior to filtration by adding chemicals such as ferric chloride and lime or organic polymers so that the sludge can be filtered at an economical rate. Another operation

which often precedes filtration is elutriation, a sludge washing operation used to lower the chemical conditioning requirements. Filtration rates for waste water sludges vary from two to fifteen pounds of dry solids per square foot of filter area per hour and final cake moisture contents vary from sixty to eighty percent (4).

Destruction of Organic Solids

Combustion

Combustion is the most practical means presently known for obtaining the maximum reduction of the organic fraction of waste water solids. Its first application to sewage solids was at Dearborn, Michigan in 1935 (18). Its acceptance in the sanitary engineering field was slow due to high capital and operating costs. Today, however, combustion is becomine economically competitive with digestion. The four major combustion processes employed for the destruction of the organic fraction of waste water solids are drying and incineration, Zimmermann combustion, fluidized bed, and atomized suspension.

Drying and incineration is the most common combustion process utilized in the sanitary engineering field. It is accomplished by air drying the sludge and feeding it to the combustion chamber of a furnace. Drying and incineration is generally preceded by mechanical dewatering which, in many cases, determines the economic feasibility of the operation. The chemicals, ferric chloride and lime, used to condition sludges for vacuum filtration can reduce the heat energy available for sustaining combustion. The final moisture content of the

dewatered sludge is the major factor which determines the amount of auxiliary fuel that will be required for the combustion process (18).

The Zimmermann combustion process is a relatively new method of oxidizing the organic fraction of waste water solids. The principle of the process is that any organic compound can be oxidized if sufficient energy is supplied to carry out the reaction. In the Zimmermann process the waste water organics are chemically oxidized in an aqueous solution by dissolved oxygen. The reactor in which the process takes place is maintained at an elevated temperature and pressure. Typical operating conditions are 500^{O}F and 1500 psig. It has been found that as much as 90 percent of the organic matter contained in waste water sludges can be oxidized by dissolved oxygen at 500^{O}F and 1200 psig. (19).

The fluidized bed method of sludge combustion consists of oxidizing the organic fraction of the sludge in a fluidized bed of hot silica. The sand is suspended by an upward movement of air. The sludge is introduced into the fluidized sand, completely mixed by the violent movement of the bed, and quickly oxidized. The violent mixing along with the reserve heat in the bed afford complete destruction of the organic solids with a minimum of excess air, approximately twenty percent (21). It is claimed that operating costs are greatly reduced by this low excess air requirement. The system is well suited for small installations as the heat retention properties of the sand bed make intermittent operations feasible. Other advantages claimed for the process are simplified operation, complete odor control, easy ash removal, and minimum space requirements (20).

The atomized suspension method of combustion consists of atomizing, drying, and oxidizing the sludge as it falls through a tower. The tower is maintained at an elevated temperature by hot gases which circulate through its walls. The sludge is atomized into small droplets (20-25μ) by a nozzle at the top of the reactor tower. As the droplets fall the moisture is evaporated and at a selected point hot air is injected to oxidize the solids. A major difficulty encountered in the use of the process is effecting atomization of the sludge (21). Advocates of the process claim that the only outside energy required is that used for pumping the liquid (17).

The processes employed for the combustion of the organic fraction of the waste water solids are often considered as methods of final disposal; however, after these processes are completed, 25-65 percent of the original dry solids must still be disposed of by some means (22). When adjacent land is available, the ash is usually dispersed in water and pumped into a lagoon. The potential air pollution problem must be considered in evaluating any combustion process.

Digestion

Sludge digestion has been the standard method employed to destroy the organic fraction of the solids in domestic waste water sludges. In this system all of the sludge solids are pumped into an enclosed air-tight tank where they decompose under anaerobic conditions. Digestion serves the dual purpose of rendering the solids readily drainable and converting a portion of the organic solids to methane. Two schools of thought prevail today in the large treatment plants. One advocates the use of sludge digestion facilities before vacuum filtration, whether

whether followed by drying or not, on the basis that there are some economies realized from gas production and less difficulty in handling the sludge. The other advocates dewatering of raw sludge followed by drying and incineration. Two methods of sludge disposal which are gaining in popularity are open only to digested sludge. These are land disposal of liquid sludge and lagooning. Digestors continue to be an important process employed in smaller treatment plants. Digestion should be investigated as a possible part of any sludge handling and disposal system.

Final Disposal

Sanitary land fills

Sanitary land filling is a relatively inexpensive method of final disposal which can be used for vacuum-filtered, centrifuged, or sand-bed-dried sludge It consists of burying the waste solids in a planned and methodical manner. The operation is usually carried out in conjunction with a municipal refuse sanitary fill project. The area used for the fill should be easily accessible, yet, remote from sources of water supply and recreational areas (23). The major cost of such an operation is the cost of the land upon which the fill is to be located.

Sludge barging

Sludge barging is an economical method of final disposal which can be utilized when land area is limited, the treatment plant is near a large body of water, and local regulations do not forbid it. In the past raw, precipitated, digested, and filtered sludges have been disposed of by barging by many of the coastline cities. The present

trend, however, is to dispose of only digested sludges by this method. Rawn and Bowermann state that past objections to this method of disposal were based on unsatisfactory disposal of raw sewage (24). New York City has barged digested sludge to the sea for years and has found the operation effective and economical (14). The sludge barging operation as usually employed consists of pumping the sludge into a barge, transporting it to a site far from shore, and discharging under the surface of the water.

Fertilizer and soil conditioner

The use of waste water sludges as fertilizers is of limited economic value in the United States because of the availability of low cost commercial fertilizers. Fertilizer production has been tried at a number of installations and has been abandoned by most of the larger plants. As usually applied, the process consists of flash drying the sludge and selling it either in bulk or bagged quantities. Fertilizer was produced at the Milwaukee and Chicago treatment plants but in both cases more economical methods of disposal have either been installed or are under consideration (8).

Most methods of final sludge disposal on the land are employed primarily to utilize the digested sludges as soil conditioners. Land disposal of liquid digested sludges holds some promise as an economic method of final disposal. The liquid sludge is hauled in tankers to designated areas and sprayed over the surface of the ground. Miami, Florida has reported disposal of three million gallons of six percent digested sludge on 35 acres in a six month period (25). The Metropolitan Sanitary District of Chicago has switched to land disposal of

liquid digested sludge (25). The Maple Lodge Works, England, found that the cheapest and most efficient method of final disposal was to apply liquid digested sludge to the land (1). New York City uses part of their digested sludge in this manner to produce an artificial top soil for city parks (14).

Whenever waste water sludges are applied to the land, the hygienic hazards involved must be considered. Pathogenic bacteria, viruses, protozoa, and worm eggs can survive digestion and air drying. The only completely safe method of land application is to apply sludge that has been heat dried (4).

CHAPTER III

VACUUM FILTRATION

Filtration can be defined as the separation of suspended, particulate solids from a fluid mixture by passage of most of the fluid through a porous medium which retains the solids on or within itself. The mixture to be separated is usually called the feed slurry; the solids which accumulate in visible amounts on the medium are called the cake; the fluid which passes through the medium is called the filtrate. The equipment which provides housing for the medium, means for the application of the vacuum, drainage surfaces, and means for the removal of appropriate streams is called the filter.

Although the exact mechanism for the arrest and accumulation of the solids is not clearly understood, it appears as if the removal is accomplished by straining, adhesion, and possibly adsorption. Two general model pictures of the filtration process have developed which provide a theory that is consistent with the majority of data collected thus far for filtration rates and resistance. The model in which the solids are stopped at the surface of the medium and build up to form a cake of increasing thickness leads to the cake-filtration equations; the model in which the solids are trapped within the pores or body of the medium leads to the filter-medium-filtration equations. The types of filtration utilized in water and waste water treatment, slow and

rapid sand filtration, micro-straining, diatomite filtration, centrifugal filtration, and vacuum filtration can be formulated by using one of these models.

To date filtration has developed mainly as a practical art rather than as a science. This is due to the many complex, interrelated parameters which affect the process; however, with the advent of the modern computer this problem should cease to be an obstacle to the designing engineer. If the objective is a least cost operation, then it is imperative that the effect of all parameters affecting the process be clearly understood.

In the cake-filtration process the result is usually a relatively clear filtrate and a layer of solid particles and liquid on the surface of the porous medium. Once the layer has formed its surface acts as the filter with a continuously increasing depth of solids being deposited on it. The head loss across the cake for a particular section of the filter is continuously increasing while the rate at which the filtrate flows through that section, thus, the rate at which the solids are being deposited, is continuously decreasing. The cake that is formed is composed of a mass of particles of irregular shapes and sizes. Through the cake run many small pores which act as small pipes for the filtrate flow.

Theoretical Development

Since vacuum filtration is a special case of flow through small pipes, the theory can be developed from the Darcy-Weisbach equation

$$h_f = f\left(\frac{L}{4R}\right)\left(\frac{v^2}{2g_c}\right), \qquad (1)$$

where h_f = frictional head loss, ft. lbf./lbm.

f = friction factor, dimensionless

L = depth of bed, ft.

R = hydraulic radius, ft.

v = average velocity (mean interstitial velocity), ft./sec.

g_c = Newtons conversion factor, ft. lbm./lbf. sec.^{2}

In order to apply equation one to flow through beds of solids, expressions must be obtained for the hydraulic radius, which is defined as cross-sectional area divided by the wetted perimeter, and the superficial velocity, which is defined as the velocity at which the fluid would flow through an empty bed. The total channel volume is equal to the volume of the voids; therefore, if

ϵ = porosity of the bed; ratio of the volume of voids to the total volume

N = total number of solid particles

V_p = volume of a single solid particle

V_v = total volume of voids

V_t = total volume of solids

the channel volume, V_v, can be expressed as

$$V_v = \left(\frac{V_v}{V_t}\right)\left(V_p\right)\left(N\right) = \left(\frac{\epsilon}{1-\epsilon}\right)\left(V_p N\right). \qquad (2)$$

The wetted surface area is equal to the number of solid particles times the surface area of a single particle, i.e.

Wetted surface area = NS_p

where S_p = surface area of a single particle.

Since cross-sectional area divided by wetted perimeter is equivalent to channel volume divided by wetted surface area, the expression for the hydraulic radius can be written as

$$R = \frac{\text{Channel volume}}{\text{wetted surface area}} = \frac{\epsilon}{1-\epsilon}\left(\frac{V_p}{S_p}\right) \quad . \tag{3}$$

If the particles of the bed are assumed to be spherical, the expression V_p/S_p can be written as

$$\frac{V_p}{S_p} = \frac{\pi D_s^3/6}{\pi D_s^2} = \frac{D_s}{6}$$

where D_s = geometric mean sieve diameter of the solid particles, ft.

For non-spherical particles the equivalent hydraulic diameter would be

$$D_h = \varphi_s D_s \tag{5}$$

where φ_s = particle shape factor, dimensionless

≤ 1

$= 1$ for a sphere.

Thus, the expression for V_p/S_p can be written as

$$\frac{V_p}{S_p} = \frac{\varphi_s D_s}{6} \quad . \tag{6}$$

A relationship between the superficial velocity and the mean interstitial velocity can be derived as follows:

Let v_s = superficial velocity, ft./sec.

Q = volumetric flow rate of the fluid through the bed, ft^3/sec.

A = cross-sectional area of bed, ft.^{2}

then $v_s = \frac{Q}{A}$, and since $v = \frac{Q}{\epsilon A}$, (7)

$$v_s A = \epsilon v A$$

and $v_s = \epsilon v$.

When equations three, six, and seven are substituted into equation one, the Carmen-Kozeny equation is obtained:

$$h_f = f' \left(\frac{L}{\varphi_s D_s}\right) \left(\frac{1-\epsilon}{\epsilon^3}\right) \left(\frac{v_s^2}{g_c}\right) \tag{8}$$

where $f' = 3f$ = friction factor, dimensionless.

The Carmen-Kozeny equation gives the head loss for a clear fluid flowing through a clean porous bed of uniformly sized particles for all regimes of saturated flow. The flow of filtrate through a filter cake is similar to the flow of a clear fluid through a bed of solid particles except that the filtrate flow causes the depth and porosity of the filter cake to change. The Carmen-Kozeny equation can be modified to suit this condition.

The flow of filtrate through a filter cake has been found to be laminar (26). A plot of the friction factor as a function of the Reynolds number for laminar flow has produced the following relationship (26):

$$f' = K' \left(\frac{1-\epsilon}{N_{RE}}\right) \tag{9}$$

where N_{RE} = Reynolds number

K' = constant, dimensionless

The Reynolds number can be expressed as

$$N_{RE} = \frac{\rho \varphi_s D_s v_s}{\mu} \tag{10}$$

where ρ = filtrate density, lbm./ft.^{3}

μ = filtrate viscosity, lbm./ft. sec.

When the superficial velocity is expressed as

$$v_s = \frac{Q}{A} , \tag{11}$$

equations nine, ten, and eleven can be substituted into equation eight to give the basic cake filtration equation:

$$h_f = K' \left[\frac{(1-\epsilon)^2}{\varphi_s^2 D_s^2 \epsilon^3} \right] \left(\frac{\mu L Q}{g_c \rho A} \right) \tag{12}$$

An examination of equation twelve shows that the head loss through a filter cake is a function of the porosity of the cake, the shape and diameter of the solid particles, the depth and surface area of the filter cake, the viscosity and density of the filtrate, and the volumetric flow of the filtrate.

When equation twelve is applied to the vacuum filtration operation, the bracketed term is equated to a parameter called the specific resistance, a characteristic property of the filter cake. The specific resistance can thus be written as

$$r = \frac{K'(1-\epsilon)^2}{\varphi_s^2 D_s^2 \epsilon^3} \tag{13}$$

where r = specific resistance of the filter cake, (ft.)$^{-2}$.

With the frictional head loss expressed as

$$h_f = \frac{\Delta P}{\rho} \tag{14}$$

where ΔP = pressure drop across the filter cake, lbf./ft.^{2},

equation (12) can be written as

$$Q = \frac{g_c \Delta PA}{\mu rL} \; . \tag{15}$$

Equation fifteen is an expression for the rate of filtration in terms of the volumetric flow of filtrate through the filter cake. The rate of filtration is thus directly proportional to the pressure drop across the filter cake and the cross-sectional area of the filter cake and inversely proportional to the viscosity of the filtrate, the specific resistance of the filter cake, and the depth of the filter cake. Equation fifteen is an expression of Darcy's Law for the laminar flow of a fluid through a porous medium when the specific resistance and depth of filter cake are assumed constant.

There are actually two resistances to the flow of filtrate in a vacuum filter, the filter cake resistance and the filter media resistance. The filter media resistance can be accounted for by an equivalent flow term and the reciprocals of the flows can be added, i.e.

$$\frac{1}{Q} = \frac{\mu rL}{g_c \Delta PA} + \frac{\mu R_r}{g_c \Delta PA} \tag{16}$$

where R_f = filter media resistance, (ft.)$^{-1}$.

Equation sixteen can be rearranged to yield the following equation:

$$Q = \frac{g_c \Delta PA}{\mu(rL+R_f)} \tag{17}$$

If the volume of filter cake deposited per unit volume of filtrate is assumed to be constant, the depth of filter cake can be expressed as

$$L = \frac{V}{A}\ c \tag{18}$$

where c = volume of solids deposited per unit volume of filtrate.

It is usually more convenient to have equation eighteen in terms of the weight of dry cake solids per unit volume of filtrate rather than the volume of solids per unit volume of filtrate; therefore, let

ω = weight of dry cake solids deposited per unit volume of filtrate, lbf./ft^3.

γ_s = unit weight of solids, lbf./ft^3.,

then c can be expressed as,

$$c = \frac{\omega}{\gamma_s}\ , \tag{19}$$

and equation eighteen can be written as

$$L = \frac{V}{A}\,\frac{\omega}{\gamma_s}\ . \tag{20}$$

Equation seventeen can be written as

$$Q = \frac{dV}{dt} = \frac{A^2 \Delta P g_c}{\mu(\omega V r' + AR_f)} \tag{21}$$

where $r' = r/\gamma_s$, ft./lbf.

t = time of filtration or cake formation, sec.

V = volume of filtrate, ft^3.

Previous investigators have assumed that the only variables in equation 21 during filtration were the volume of filtrate, V, and the time of filtration, t. By separating these variables equation 21 can be integrated to yield the time of filtration:

$$\int_o^t dt = \int_o^V \left(\frac{\mu\omega Vr}{A^2 \Delta P g_c} + \frac{\mu R_f}{A \Delta P g_c} \right) dV \tag{22}$$

$$t = \left(\frac{\mu\omega r'}{A^2 \Delta P g_c} \right) \frac{V^2}{2} + \left(\frac{\mu R_f}{A \Delta P g_c} \right) V \tag{23}$$

Equation 23 can be rearranged in the following manner:

$$\frac{2t\Delta P g_c}{\mu\omega r'} = \frac{V^2}{A^2} + \frac{2R_f V}{\omega r' A} \tag{24}$$

$$\frac{V}{A} = \left(\frac{2\Delta P t g_c}{\mu\omega r'} - \frac{2R_f V}{\omega r' A} \right)^{1/2} \tag{25}$$

Equation 25 is an expression for the volume of filtrate obtained per unit area of filter during time t. It is converted to the vacuum filter yield equation in the following manner:

Let Y = filter yield, lbf./ft.^{2} sec.

t_c = filter cycle time; time for one revolution filter drum, sec.

then

$$Y = \frac{\omega V}{t_c A} = \left(\frac{2\Delta P \omega t g_c}{\mu r' t_c^2} - \frac{2\omega R_f V}{r' A t_c^2} \right)^{1/2} \tag{26}$$

The relationship between the filter cycle time and the time of cake formation can be expressed as,

$$t = \alpha t_c \tag{27}$$

where α = fraction of cycle time occupied by cake formation. Thus, the filter yield can be expressed as

$$Y = \left(\frac{2\Delta P \omega \alpha g_c}{\mu r' t_c} - \frac{2\omega R_f V}{r' A t_c^2} \right)^{1/2} \qquad (28)$$

Equation 28 is a general form of the most widely used equation to describe the vacuum filtration operation (26).

The theory as presented in this paper is basically Carman's theory; however, it is developed in a manner that shows the parameters which affect the value of the specific resistance, r (26). Others who have contributed to the theoretical approach to vacuum filtration are:

Hatschek	1908, first to develop theory (27).
Ruth & Carman	1933, established a parabolic relationship between volume of filtrate and time of filtration and introduced the concept of "average specific resistance" (28) (29).
Thompson & Proctor	1938, investigated Carman's work using Buchner funnel (30).
Coackley & Jones	1956, investigated Carman's theory and found the concept of average specific resistance to be of value in expressing filtration results (31).
Hulff	1952, attempted to describe filtration by analogy with heat transfer theory, method difficult to apply (32).
Grace	1953, obtained values for average specific resistance from compression - permeability data, method is time consuming and requires complex apparatus (33).
Beck, Sakellarioi, & Krup	1955, modified Buchner funnel, test to evaluate parameters (34).
Jones	1956, modified Carman's theory and used concept of average specific resistence to predict filter yield (35).

Investigation of Parameters

Laminar flow

The only assumption made in obtaining the basic cake filtration equation was that laminar flow conditions existed for the flow of filtrate. It is unlikely that the initial flow of filtrate through the filter medium is laminar; however, for the major portion of the cake formation time the flow is through the filter cake and has been found to be laminar (26).

Constant deposition of cake solids

The assumption that a constant volume of cake is deposited per unit volume of filtrate is good for fine filter media, such as woven cloth media which allow only approximately one percent of the solid particles to pass through with the filtrate; however, it is not true for coarse filter media, such as the metallic media. A significant portion of the sludge solids will pass through a coarse filter medium with the initial flow of filtrate and thus, invalidate the assumption. After the filter cake has formed, equal volumes of filtrate will again deposit equal volumes of cake solids. This parameter will be investigated further in Chapter Four.

Specific resistance

The specific resistance of the filter cake is a function of the cake porosity and the size and shape of the solid particles in the

filter cake, i.e.,

$$r = K' \left[\frac{(1-\epsilon)^2}{\varphi_s^2 D_s^2 \epsilon^3}\right] \tag{13}$$

Previous investigators using Carman's theory have assumed that a constant resistance is added for equal volumes of filtrate. This assumption leads to the concept of an "average specific resistance" for the filter cake. It will be shown in Chapter Four that this assumption is approximately true for fine filter media; however, the assumption requires further investigation for coarse filter media. When a coarse filter medium is used, many of the fine solid particles will pass through the filter with the initial flow of filtrate leaving a relatively porous filter cake with a low resistance to flow. After the cake has reached a certain thickness, essentially all of the solid particles in the sludge will be stopped in the cake and the porosity of the cake will assume a constant value characteristic of the sludge being filtered. The specific resistance will then remain constant for the remainder of the cake formation time, i.e. the incremental increase in the head loss per unit volume of filtrate filtered will be constant.

Carman has shown that the average specific resistance of many filter cakes is a function of the applied pressure differential and can be expressed as

$$r = r_o(\Delta P)^n \ . \tag{29}$$

where r_o and n = constants (29).

A filter cake with a resistance to the flow of filtrate that follows this relationship is called a compressible cake, and n is known as the coefficient of compressibility. When $n = 0$, $r = r_o$ and the cake is incompressible. Experimentation has shown that this relationship holds for all types of waste water sludges (32). Values obtained by Coackley and Jones are shown in table 3-1 (31).

TABLE 3-1

TYPICAL VALUES OF COMPRESSIBILITY FOR SEWAGE SLUDGE

Type of Sludge	Range in value of n
Digested	0.70 to 0.86
Activated	0.60 to 0.79
Raw	0.87 (one sample)
Humus	0.80 (two samples)

Compressibility can be accounted for in equation thirteen by the shape factor term. As non-rigid particles are compressed they become less spherical; thus, the shape factor decreases and the specific resistance increases.

Application of Theory

There have been few attempts to predict full-scale performance from laboratory determinations of average specific resistance values. The only published data found of such an attempt was that by Jones (35). He calculated the average specific resistance as previously described and, by assuming the filter medium resistance was zero, used the

following modified yield equation to predict filter yields:

$$Y = 0.0357 \; \frac{100-C_f}{C_i-C_f} \left[\frac{m\Delta P C_i (100-C_i)}{R_r \mu t_c} \right]^{1/2} \qquad (30)$$

where Y = filter yield, lbs dry solids/ft^2/hr.

C_f = final moisture content of cake, percent

C_i = initial moisture content of the sludge, percent

m = fraction of time which suction acts, percent

ΔP = average suction pressure, lbf./sq.in.

t_c = time for one revolution of the drum, min.

R_r = specific resistance of the sludge times 10^7, sec^2/g.

μ = viscosity of filtrate, cp.

The results of his work are summarized below:

TABLE 3-2

COMPARISON OF PREDICTED AND MEASURED FILTER YIELDS

Type of Sludge	Predicted Yield lbs. dry solids/ft/hr	Measured Yield lbs. dry solids/ft/hr
Mixture-elutriated, digested and activated	1.25	1.07
Conditioned activated	1.43	1.27
Conditioned digested	7.58	7.84

The type of filter medium used on the filters in these tests was not presented; however, it is reasonable to assume that it was some type of cloth medium. The results of these tests support the theoretical model and the use of an average specific resistance value in

the filter yield equation.

Operational Variables

Filter media

The filter medium used for vacuum filtration plays an important part in determining if the desired filtration objective is to be achieved. It affects the quality of the filtrate, the filtration rate, the final moisture content of the filter cake, and the chemicals required for conditioning.

The media used on the early vacuum filters were usually some type of woven cloth. These media acted as fine strainers which resulted in high resistances to the flow of filtrate. Yields of one to two pounds of dry solids per unit area per hour and suspended solids concentrations in the filtrate of 100 mg/l were common (36). These media were short lived, clogged easily, and required a large amount of maintenance work.

The trend today in the vacuum filtration of waste water solids is to use the coarse weave synthetic media or the metallic media. The use of very coarse filter media has enhanced the economics of vacuum filtration and helped to make it competitive with the other methods of dewatering sludges. Coarse filter media allow a relatively high percentage of the fine solid particles to pass through the filter with the filtrate; this fact increases filtration rates and decreases filtration costs. The solids which are recycled back to the treatment plant create a small additional solids load. At Minneapolis, St. Paul, the average vacuum filter recycled solids load is equivalent to five per-

cent of the total suspended solids reduction through both primary and secondary treatment (8). The high suspended solids concentration in the filtrate is considered necessary for higher filtration rates and when designing their new facility, this load was included as part of the total suspended solids load for the primary sedimentation tank. At Boulder it was found that the recycled, conditioned filtrate increased the suspended solids removal efficiency of the treatment plant.

Table 3-3 shows a comparison of two filter media of different pore diameter which was obtained at a primary treatment plant (37).

TABLE 3-3

EFFECT OF SIZE OF OPENING ON VACUUM FILTRATION PERFORMANCE

	120x46 mesh satin finish monofilament	S/S Spring .018" nominal opening
Yield, lbs/ft^2/hr	2.1	7.6
Conditioning required:		
% CaO used	16.0	8.8
%FeCl used	6.7	1.4
% Moisture in cake	70.9	74.5

The effect of filter medium pore diameter on filtration rates will be examined further in Chapter Four.

Sludge characteristics

The filtration rates for vacuum filters using the coarse media are dependent entirely upon the characteristics of the solids which form the cake, as it is the cake that is performing the filtration. This

is true provided the filter has the required liquid and solids discharge capabilities. The characteristics of the sludge which affect the filtration operation are:

(1) particle size and shape

(2) viscosity of the filtrate

(3) initial solids concentration

(4) compressibility of sludge particles

(5) chemical composition and particle charge

The means available for altering these characteristics are conditioning, elutriation, and thickening.

Particle size and shape. The size and shape of the solid particles will affect the value of the specific resistance as shown by equation (13). These characteristics will also influence the porosity of the filter cake, as small irregularly shaped particles will form a more compact cake than large, regularly shaped particles. The size of the particles is the major factor determining the compressibility of the filter cake. It has been found that small particles produce relatively high compressibilities (38). At Hyperion it was found, after prolonged elutriation had washed the fine material from the sludge leaving a "chaff-like" material, that filtration yields were 30 percent greater using only 80 percent of the original conditioning chemicals (38). Thermophilic digestion, which produces larger sludge particles than mesophilic digestion, gave a three-fold increase in filter yield with a 50 percent reduction in conditioning chemicals (3). The filtration rates for the different types of waste water sludges, as shown in

table 3-4, can be related to the size of the solid particles in these sludges (3).

TABLE 3-4

MINIMUM FILTER RATES AFTER CHEMICAL CONDITIONING FOR NON-CLOGGING FILTER MEDIA

Type of Sludge	Yield lbs. dry solids/ft^2/hr Fresh	Digested
Primary	8	7
Primary & trickling filter	7	6
Primary & activated	6	5
Activated or trickling filter	3	-

Viscosity of filtrate. The rate of filtration varies inversely with the viscosity of the filtrate; thus, if the temperature of the sludge could be raised, the viscosity would be lowered and filtration rates would be increased. Since heating of waste water sludge is uneconomical, the viscosity is of little importance in effecting an increase in filtration rates.

Initial solids concentration. The weight of dry cake solids deposited per unit volume of filtrate can be related to the solids concentration of the unfiltered sludge in the following manner (26):

$$i = \frac{\omega V}{\left[(z\omega V - \omega V)g_c/\rho g\right] + V} \qquad (31)$$

where i = initial solids concentration

z = ratio of wet cake to dry cake weight

It can be seen that the filtration rate varies as the initial solids concentration. If the solids concentration of a sludge is doubled, only half of the original filtrate must be removed to deposit the same quantity of solids; thus, the capital investment for the filter is reduced. Experiments conducted by Coakley and Jones indicated that the average specific resistance varied with the initial solids concentration (31); however, equation 13 shows they are independent. A possible cause for this finding could have been the passage of more fine solid particles with the dilute sludge than with the concentrated sludge.

Compressibility. The term compressibility as it applies to filter cakes describes in qualitative terms the tendency of the solid particles in the cake to deform. When the cake solids are compressible, they will form a cake in which the specific resistance will vary as the applied pressure differential. One method of determining if the cake is compressible is to make average specific resistance determinations at various vacuum levels. Another basis on which to judge a cake's compressibility characteristics is the slope of the line relating Log Y to Log ΔP. If it is less than 0.5, the cake is compressible. This criteria can be derived from equations 28 and 29. If a filter cake is compressible, diminishing returns on yield will be obtained for each incremental increase in the cake

formation vacuum level.

Chemical composition and partial charge. The chemical composition of the sludge and the charge on the solid particles are important parameters to be considered in the destabilization of the fine solid particles in order to form larger particles. Sewage sludges are complex colloidal systems of large organic molecules. The particles are generally hydrophilic in nature and possess negative charges on their surfaces due to the ionization of certain functional groups or the preferential adsorption of specific ions. The stability of the system is a function of the bound water layer and the charge on the particles. These properties are affected by the pH and ionic strength of the dispersing fluid (39).

Sludge conditioning. Sludge conditioning is a means of improving the filterability of a sludge by the addition of chemicals which alter the characteristics of the sludge solids. The objective of the process is to induce the fine solid particles to agglomerate in some way to form larger particles which will form a pervious filter cake and thus, produce higher filtration rates. The process involves chemical reactions of inorganic compounds, a reduction or reversal of the zeta potential, coagulation, and flocculation. Genter has shown that when inorganic conditioning chemicals are used the amount of chemical required is a function of the bicarbonate alkalinity of the liquid and the volatile fraction of the suspended solids (40). The following reactions can occur in the liquid phase when ferric chloride and lime are used:

$$2FeCl_2 + 3Ca(HCO_3)_2 \rightarrow 2Fe(OH)_3 + 3CaCl_2 + 6CO_2$$

$$NH_4HCO_3 + Ca(OH)_2 \rightarrow CaCO_3 + 2H_2O + NH_3$$

The primary function of these conditioning chemicals is to lower the zeta potential so that the particles will coagulate.

One of the factors which has improved the economic feasibility of vacuum filtration has been the introduction of organic conditioning chemicals (39). These chemicals are water soluble long chain organic compounds of high molecular weight called polymers. In solution they ionize to form long chain molecules with many positive or negative sites. Those which form positive sites are called cationic, and those which form negative sites are called anionic (41). Cationic polymers are used most widely in sewage treatment. The primary mechanisms involved in the use of polymers are particle adsorption and "bridging" by the long chain molecules. This process is commonly referred to as flocculation (42). Dispersions which have been flocculated by polymers yield fluffy 3-dimensional flocs which possess large pores. These pores permit rapid filtration through the filter cake. It has been found that the filtration rate is dependent upon the eighth power of the concentration of the polymer (42). This law is quite different than that obeyed by simple salts acting as coagulants; "The efficacy of a given salt depends primarily upon the valency, i.e. roughly to the sixth power of the charge..."(42).

Elutriation. Elutriation is a solids-washing process generally employed to reduce the bicarbonate alkalinity of digested sludges

and therefore, lower the so called "liquid demand" for the coagulent (40). As mentioned previously, the removal of fines is also an important factor in this process. Operators have reported that elutriation partially solves one problem by shifting it to another point in the treatment process where it must be handled in another way (3). Elutriation is generally employed when digested sludges are being filtered because they are relatively high in bicarbonate alkalinity and fine particles. Genter has developed the following formulas to determine the alkalinity remaining in the sludge liquid after multi-stage and two stage countercurrent elutriation (40).

for multi-stage, $$E = \frac{D + W\left[(B+1)^{s}-1\right]}{(B+1)^{s}} \qquad (32)$$

and for countercurrent, $$E = \frac{D + (B^{2}+B)W}{B^{2} + B + 1} \qquad (33)$$

where E = alkalinity in the elutriated sludge, ppm

D = original alkalinity, ppm

B = ratio of volumes of elutriating water to the volume of liquid in the sludge

s = number of stages employed

W = alkalinity of the elutriating water, ppm

Vacuum

When R_f is set equal to zero and equations 28 and 29 are combined, the relationship between filter yeild and pressure drop across the cake can be written as

$$Y \approx \Delta P^{(1-n)/2}$$

This equation shows that the yield varies as the applied pressure differential when n is less than one. Jones has found that most sewage sludges treated with ferric chloride have a compressibility coefficient of 0.85 and that for these sludges little advantage is gained by increasing the vacuum level (35) Schepman and Cornell, however, have concluded that the highest possible vacuum should be used during cake formation since for any value of n less than one, an increase in vacuum will produce an increase in yield (43). They point out that the cost of a vacuum pump is more a function of the air rate than the vacuum level and that essentially no air is pulled by the pump during cake formation.

Cake thickness

Although cake thickness neither appears as a variable in the yield equation, or is normally considered as such, an examination of the derivation of the equation will show that the yield is related to the cake thickness. Each incremental layer of solids deposited on the cake creates an incremental increase in the total resistance to the flow of filtrate; thus, it can be concluded that ideally the cake should be as thin as possible. The flow of filtrate obtained from a continuous filtration operation will be inversely proportional to the cake thickness maintained. The actual relationship will be dependent on the rate of increase of the specific resistance of the sludge being filtered. The cake thickness can be controlled by varying the drum speed. Common cake thicknesses range from 1/16 to 1/4 inch.

Cycle time

The filter cycle time for a rotary drum vacuum filter is the time required for one complete revolution of the filter drum. It is divided into three phases, which are cake formation, cake dewatering and cake discharge, see figure 3-1. The cake formation phase is that portion of the cycle time in which the drum is submerged in the sludge and filtration is taking place. The cake dewatering phase is the time between drum submergence and cake discharge during which a vacuum is applied to reduce the moisture content of the filter cake. The cake discharge phase is that portion of the cycle time required to dislodge the filter cake from the filter media so that the cycle can begin anew. An examination of equation 28 shows that the yield is inversely proportioned to the square root of the cycle time; however, there is a mechanical limit on the benefits to be gained in this way. This limitation is imposed by the minimum thickness of cake that can be successfully discharged from the filter. An examination of the derivation of equation 28 shows that the filtration rate is directly proportioned to the square root of the cake formation time; however, there is also a limit imposed on this relationship. The initial increment of cake thickness for a coarse medium will form a relatively porous cake. As filtration continues, fine particles as well as coarse particles will be added to the cake causing the resistance to the flow of filtrate to increase. This resistance will cause the head loss across the cake to approach the available head; thus, the flow of filtrate will approach zero. The fraction of the cycle time alloted to each phase can be varied by altering drum speed and

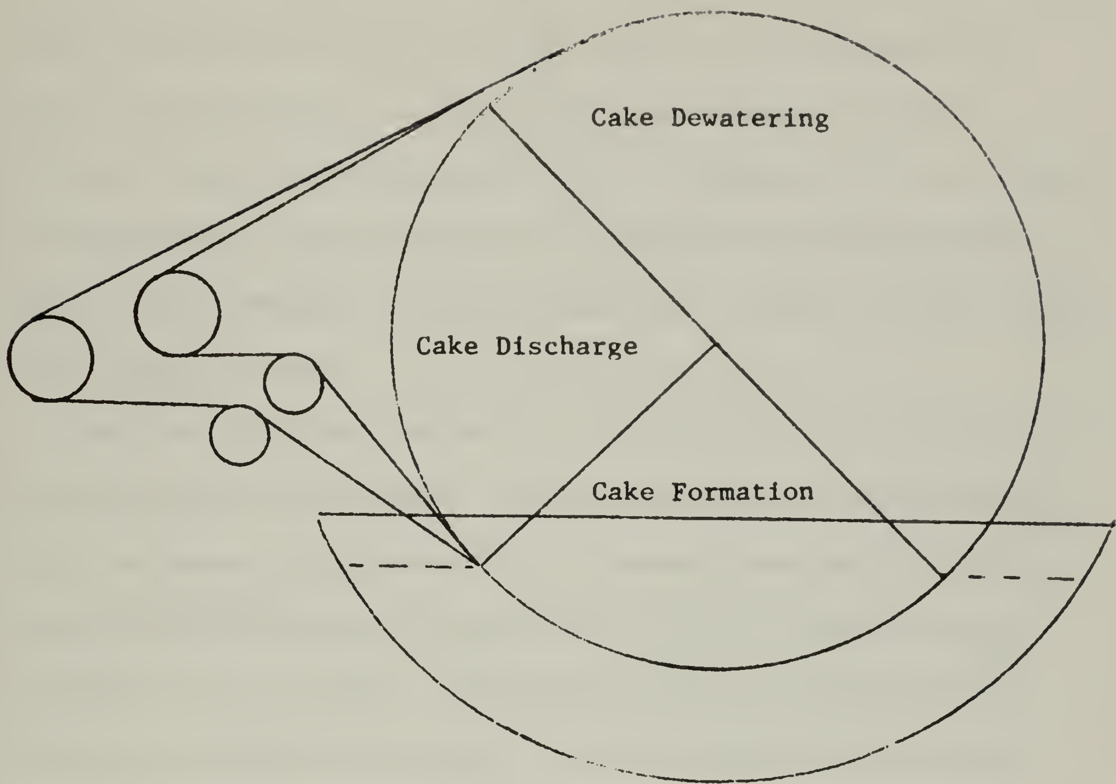

Figure 3-1. Vacuum Filter Cycle

submergence. A typical division of the cycle time is 25% cake formation, 50% cake dewatering and 25% cake discharge.

Filter cake moisture

The three criteria which are usually used to evaluate the performance of a vacuum filter are filter yield, which is basically the same as filtration rate, final cake moisture, and concentration of suspended solids in the filtrate. Yield has been discussed in some detail in previous sections and the suspended solids in the filtrate are normally not a problem; however, the final moisture content of the filter cake is often a very important factor to be considered. The importance will

depend on the method of final solids disposal employed. If the solids disposal is to be by incineration, the objective will be a burnable cake of the lowest possible moisture content so as to reduce the costs of auxiliary fuel. If the solids are disposed of by depositing them as a sanitary land fill, the objective will be a final moisture content which produces a sludge which can be handled and hauled. The economical degree of dewatering in this case will depend on the length of haul.

The final moisture content of a filter cake is dependent on the characteristics of the sludge, the filter media, and cake dewatering time. The amount of moisture that is removed from the filter cake during cake dewatering will be determined by how effectively the cake pores can be drained. The lowest practical moisture content that can be obtained occurs when all of the cake pores have been drained. The most effective cake draining will be achieved with a filter cake composed of uniformly sized, regularly shaped particles. This type of cake will have pores that will drain at approximately the same rate. Cakes composed of nonuniformly sized, irregularly shaped particles will have some pores which will drain much faster than others. This situation will produce a loss of vacuum before all the pores in the cake have been drained, and thus, a higher final moisture content. When raw, primary sludge is filtered at the Boulder treatment plant on a "Coilfilter", the vacuum level holds at approximately five inches of Mercury; however, when digested sludge is mixed with the raw sludge, the vacuum will hold at approximately twelve inches of Mercury and a lower final cake moisture will be obtained.

It was found that the cake dewatering of the raw primary sludge at the Boulder plant reduced the moisture content of the cake by approximately five percent.

Parameter Evaluation by Previous Investigators

The most common method of determining the filterability characteristics of waste water sludges is the Buchner funnel test. This test is run by filtering a volume of sludge through a filter paper with a constant vacuum and is primarily used to determine the average specific resistance and the compressibility of a sludge. Equation 23 can be written in the following manner:

$$\frac{t}{V} = \left[\frac{\mu\omega r'}{2A^2\Delta P g_c}\right] V + \frac{\mu R_f}{A\Delta P g_c} \qquad (23)$$

This equation is a linear expression with t/V and V as the variables; thus, by filtering a sample of sludge and recording the volume of filtrate at various time intervals, a graph of t/V as a function of V can be plotted. The slope of the straight line of best fit can be used to compute the average specific resistance,

$$\text{Slope} = \frac{\mu\omega r'}{2A^2\Delta P g_c} \qquad (34)$$

and the intercept can be used to compute the resistance of the filter medium,

$$\text{Intercept} = \frac{\mu R_f}{A\Delta P g_c} \qquad (35)$$

Equation 29 can be written in the following manner:

$$\text{Log } r' = \text{Log } r_o + n\text{Log } \Delta P \qquad (29)$$

This equation is a linear expression with $\log r'$ and $\log \Delta P$ as variables and a slope equal to the compressibility coefficient of the sludge. The average specific resistance can be computed at different vacuum levels and a plot of $\log r'$ as a function of $\log \Delta P$ will give the value of n. The Buchner funnel is often used to obtain a measure of the filterability of a sludge by merely noting the time until the cake cracks. This test provides the operator with an approximate idea of what type of conditioning will produce the best results on the filter.

The results obtained from a Buchner funnel test as it is usually run are dependent upon variables which are not fully defined in the determination. The initial high rate of filtrate flow is not taken into consideration in this test. Such factors as initial solids concentration, filter media, suspended solids in the filtrate, and volume of sample will have a definite influence on the results. "The results from a standard Buchner funnel test are not directly applicable to full scale operation and can only serve the purpose of arbitrarily comparing the treatments which any one worker has used" (31).

Another method of investigating the variables which affect the filtration of waste water sludges is the leaf filter test. This test can be conducted so that it actually simulates the vacuum filter operation in cake formation, cake dewatering, and cake discharge times. The results obtained are generally representative of what can be achieved from plant scale operations. The yield from a filter can be computed from the equation

$$Y = \frac{\omega V}{t_c A} \tag{36}$$

Neglecting the resistance of the filter the theoretical effects of vacuum and cycle time and submergence can be investigated with the equation

$$Y = C' \left[\frac{\Delta P^{1-n}}{t_c} \right]^{1/2} \qquad (37)$$

where C' = a constant.

The effect of cake formation time for a given sludge and a constant cycle time can be investigated with the equation

$$Y = C'' \left[\Delta P^{1-n} (t) \right]^{1/2} \qquad (38)$$

where C'' = a constant.

The leaf filter test has the advantage over the Buchner funnel test in that it simulates the actual filter operation.

CHAPTER IV

EXPERIMENTAL METHODS AND RESULTS

The objectives of vacuum filter studies have been to establish means for increasing filtration rates and decreasing costs. One of the important control variables of vacuum filtration that has not been evaluated is the affect of the size opening of the filter medium in increasing filtration rates. By increasing the size of the filter medium openings some of the fine solids of the sludge will be allowed to pass through the filter with the filtrate and will not be held in the filter cake pore openings. It would seem that the retention of the fine solid particles in the filter cake pore openings would account for a major part of the specific resistance of the filter cake. The use of filter media which allow the fine solid particles to pass with the filtrate should, thus, result in lower specific cake resistance values during the early stages of cake formation; therefore, lower average specific values and greater yields should be obtained for the filter cake.

If the size of the openings of a filter medium were small enough to capture all of the suspended solids, the specific resistance of the filter cake would be both a constant and a characteristic parameter of the sludge; however, varying degrees of solids capture by filter media produce varying specific resistance values for a sludge. The standard vacuum filtration tests as described in Chapter Three

generally employ filter media that capture essentially all of the suspended solids. The results from these tests indicate a characteristic, constant increase in flow resistance per unit volume of filtrate, i.e., a constant specific resistance. Modern vacuum filters, however, using coarse filter media, with relatively large pore openings, do not operate in this manner. These filters allow a certain fraction, five to ten percent, of the solids to pass through the filter with the filtrate and, thus, produce filter cakes with variable specific resistance values which are dependent on the filter medium pore size, depth of filter cake, and sludge particle size. The specific resistance of the filter cakes for these filters will increase as the cake builds up and begins to capture more of the fine solid particles. Thus, the average specific resistance for these filter cakes will be lower than indicated by the Buchner Funnel test and yields, since they are based on average specific values, will be greater.

Objective

The objective of the laboratory work was to investigate a variable specific resistance by a controlled filtration operation which would allow the specific resistance to be determined for each incremental layer of cake solids deposited on the filter media. With this operation the parameters affecting the filtrate flow characteristics of the filter cake were investigated in detail. Filter media with various pore diameters were employed to investigate their effect on the specific resistance and how the passage of different fractions

of sludge solids could be used to produce higher filtration rates. The initial feed solids concentration and the form vacuum were varied to investigate their influence on the rate of filtration.

Apparatus

Figure 4-1 is an illustrative sketch of the apparatus which was used in the laboratory investigation. It was constructed so that the applied vacuum, filtrate with time, and filtrate suspended solids could be measured. The vacuum application valve permitted a nearly instantaneous application and termination of vacuum to the filter medium. Different sized rounded sands placed over the perforated filter surface of the Buchner funnel served as the filter medium.

Procedure

The filtration tests were performed by placing a one-half-inch layer of washed sand uniformly over a one-fourth inch layer of fine gravel covering the perforated surface of the six-inch diameter Buchner funnel. With the vacuum application valve closed, the system was exhausted and the vacuum level adjusted with the vacuum control valve. The space between the vacuum application valve and the filter medium surface was filled with water and maintained in this condition throughout the test. An incremental volume of raw primary sewage sludge, 100 ml., was uniformly distributed over the surface of the filter medium and filtered by opening the vacuum application valve until the filter cake had formed, i.e., the liquid level was again at the filter medium surface. With the vacuum removed a small amount of fine gravel was placed over the

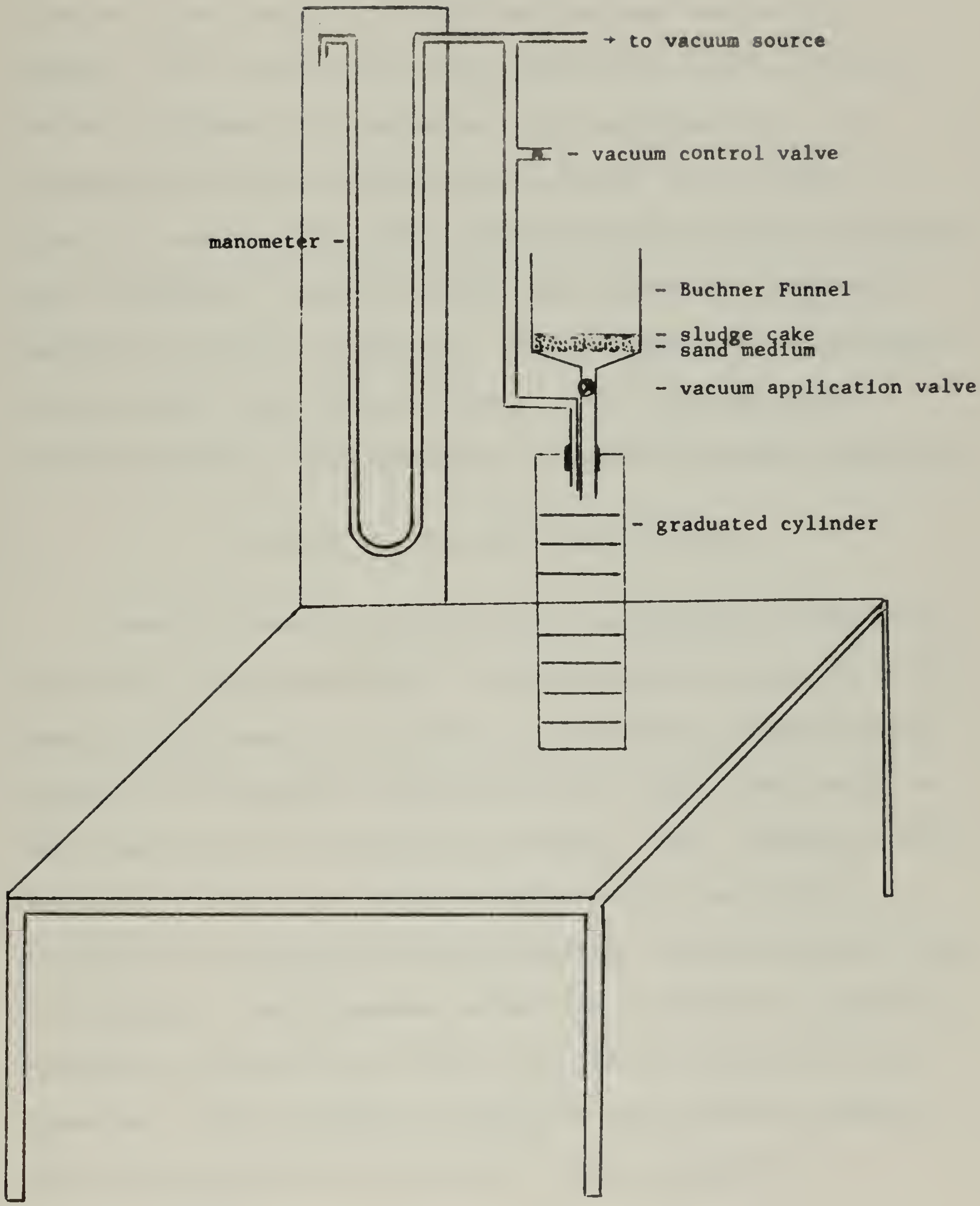

Figure 4-1. Experimental Apparatus

filter cake. A fine screen was placed over the fine gravel and a measured amount of tap water was carefully added. After removing the screen, the vacuum application valve was opened and the volume of water pulled through the filter and the time required were recorded. The filtrate water thus obtained was removed from the graduated cylinder for a suspended solids determination. The succeeding increments of sludge were added and the described procedure repeated. The screen and fine gravel were used to prevent cake disturbance. The gravel was of such a size as to present a negligable additional resistance. The measurement of the flow rate of clear water after each increment of cake formed was used to establish hydraulic characteristics indicative of specific resistance.

Analysis of Laboratory Investigation

A familiar method of presenting the experimental results would have been to plot permeability of filter cake, the reciprocal of the specific resistance, as a function of the volume of sludge filtered; however, the incremental cake depths were of such a small magnitude that they could not be accurately measured. Thus, a parameter which would not require a cake depth determination but would still illustrate the changing flow characteristics through the filter cake was developed. The resistance to the flow of filtrate in the filter apparatus is created by the filter cake, filter medium, and filter apparatus. When the medium and apparatus resistances are combined, the total resistance to filtrate flow can be expressed as

$$R_T = R_c + R_a \tag{39}$$

where R_T = total resistance to flow, sec.

R_c = cake resistance to flow, sec.

R_a = medium and apparatus resistance to flow, sec.

Equation 15 can be combined with equation 14 to yield

$$Q = \frac{h_f A}{(\mu r L/g_c \rho)} \tag{15A}$$

The flow, when expressed in this manner, can be considered as being directly proportional to a driving force, $h_f A$, and inversely proportional to the resistance, $\mu r L/g_c \rho$. Thus, the flow can be expressed in the following manner:

$$Q = \frac{h_f A}{\text{Resistance}} \tag{40}$$

or by rearranging, the resistance to flow can be written as

$$\text{Resistance} = \frac{h_f A}{Q} \tag{41}$$

Substitution of equation 41 into equation 39 yields the following equation:

$$\frac{h_f A}{Q_T} = \frac{h_f A}{Q_c} + \frac{h_f A}{Q_a} \tag{42}$$

where Q_T = total flow, ft.^{3}/sec.

Q_c = equivalent flow through filter cake, ft.^{3}/sec.

Q_a = equivalent flow through filter medium and apparatus, ft.^{3}/sec.

Since the quantity $h_f A$ is a constant for each series of tests, equation 42 can be written as

$$\frac{1}{Q_T} = \frac{1}{Q_c} + \frac{1}{Q_a} \tag{43}$$

and

$$Q_c = \frac{Q_a Q_T}{Q_a + Q_T} \tag{44}$$

The equivalent flow through the filter media and apparatus, Q_a, can be determined prior to the addition of any sludge and will remain constant during cake buildup. The total flow, Q_T, can be determined directly from the experimental data (see appendix A). The equivalent flow through the filter cake, Q_c, represents the changing flow characteristics of the filter cake and the total flow, Q_T, represents the actual flow characteristics of the filter cake, medium, and apparatus combined. Since the volume of filtrate, V, can be related directly to the volume of sludge filtered, V_s, it can be seen that, if the specific resistance and the volume of solids deposited per unit volume of sludge are constant, a hyperbolic relationship will exist between Q_c and V_s. An examination of the derivation of equation 44 will also reveal that Q_c and Q_T are inversely related respectively to R_c and R_T through the factor $h_f A$; thus these parameters, Q_c and Q_T, can be used to examine the hydraulic characteristics indicative of the specific resistance of the filter cake.

Results and Discussion

The total flow of filtrate through the filter cake, apparatus, and medium after each incremental addition of sludge for four different mesh size filter media is presented in Figure 4-2. Since each 100 cubic centimeters of sludge increased the cake depth by approximately one millimeter, these results could be interpreted in terms of an equivalent cake depth as well as the volume of sludge filtered. The results as presented in Figure 4-2 clearly show the effect of the different media on the flow rate as the filter cake is being formed.

The total flow as determined prior to the addition of sludge, i.e., at $V_s = 0$, represents the value of Q_a for each of the filter media. It can be noted that Q_a was essentially the same for the 4-9, 10-14, and 20-30 mesh sands. For these filter media the initial resistance to filtrate flow was created almost entirely by the valve and tubing of the apparatus; thus, Q_a was nearly independent of the filter media. The smaller pore diameter filter media, the 60-80 mesh sand, added significantly to the initial resistance to flow and caused a reduction in Q_a.

The coarse sand filter media produced significantly higher filtration rates than did the fine, 60-80 mesh, sand medium. The higher total flow rates exhibited by the filter cakes formed on the coarse sand filter media were created by the passage of many of the

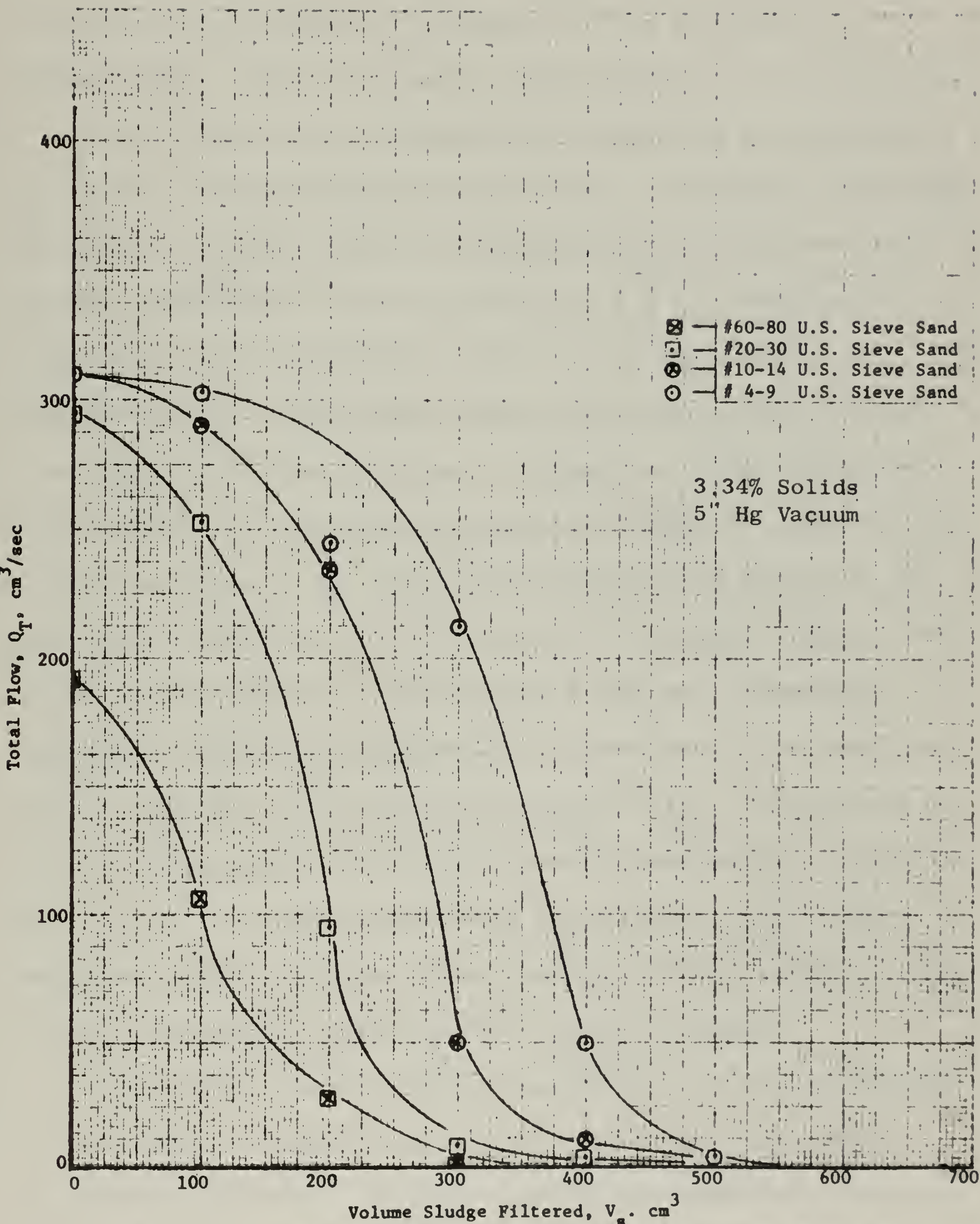

Figure 4-2. Total Flow as a Function of the Volume of Sludge Filtered.

fine sludge solid particles through the pore openings of the filter cake and medium. Since the fine sand medium trapped more of the fine sludge particles at the medium surface, the resistance, which is inversely proportional to the square of the particle size, increased very rapidly and caused a rapid reduction in filtrate flow.

If the flow rates as represented by Figure 4-2 are compared at one value of sludge filtered, such as 300 ml, a remarkable difference can be seen. Since the three large sized sand media had the same initial resistance to flow as indicated by Q_a, the difference in flow rates can be attributed to variations in the specific resistance of the filter cakes caused by the varying fractions of fine sludge solid particle passage through the filter medium with the filtrate. As succeeding increments of sludge were added to the filter cake, more of the fine sludge particles were stopped by the previously formed cake, and the flow rate was rapidly reduced. The filter cake resistance to flow created by the last incremental addition of sludge was approximately the same for all the media as shown by the small and nearly equal values of Q_T. At this state of cake formation nearly all of the fine solids were held in the filter cake instead of passing through with the filtrate. The filtrate suspended solids at this point were similar to that obtained for the initial increments of sludge on the fine filter medium.

The percent of applied solids passing the filter as a function of the volume of sludge filtered for a coarse and a fine medium is shown in Figure 4-3. In effect, the filter medium was acting as a solids classifier and during the initial period of cake forma-

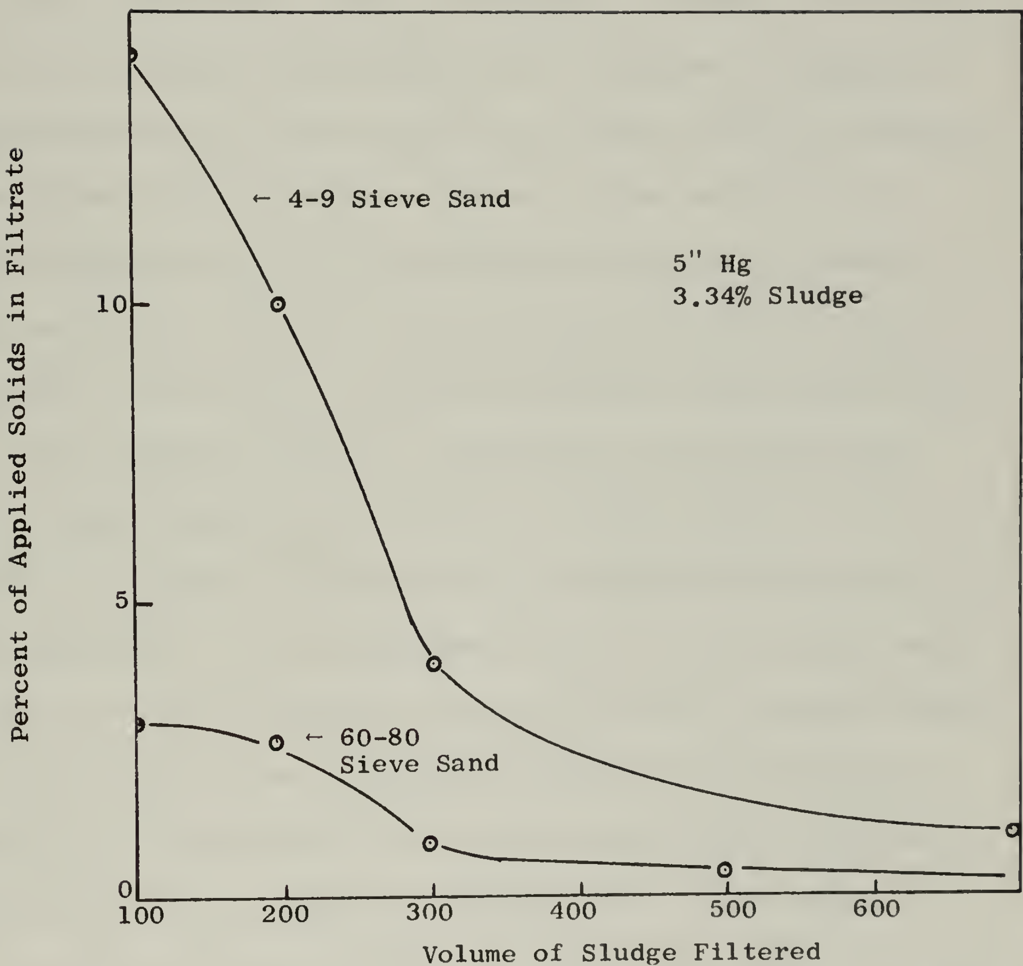

Figure 4-3. Percent of Applied Solids in Filtrate as a Function of the Volume of Sludge Filtered.

tion high filtration rates were accomplished at the expense of and by the passage of the fine sludge particles through the filter medium with the filtrate.

A more definitive illustration of the changing filter cake specific resistance as a function of the volume of sludge filtered is shown in Figure 4-4. In this figure the filtration rate without the effect of the filter medium or apparatus, is shown. The values of Q_c were calculated from equation 44 as indicated in Appendix A. A logarithmic scale has been used for the plots of Q_c as a function of the volume of sludge filtered because of the wide range of filtrate flow rates. All of the curves initiate at an infinite flow rate for zero volume of sludge filtered or zero thickness of filter cake. If none of the sludge solids were passed through the filter cake with the filtrate, a hyperbolic relationship would result. This condition would result in a curve similar to and slightly to the left of the curve for the sludge filtered on the 60-80 mesh medium. It can be assumed that all of the curves shown in Figure 4-4 would converge if they were carried to larger volumes of sludge filtered. The increased filtration rates produced by the passage of fine solids in the early stages of cake formation for the larger filter media are clearly shown by this figure.

The curves plotted in Figures 4-5 and 4-6 are similar to those described in Figures 4-2 and 4-4; however, they present the flow rates for the 20-30 mesh filter medium at various vacuum levels. These results clearly show that for a given filter medium and sludge the highest vacuum level produces the highest filtration rate. It

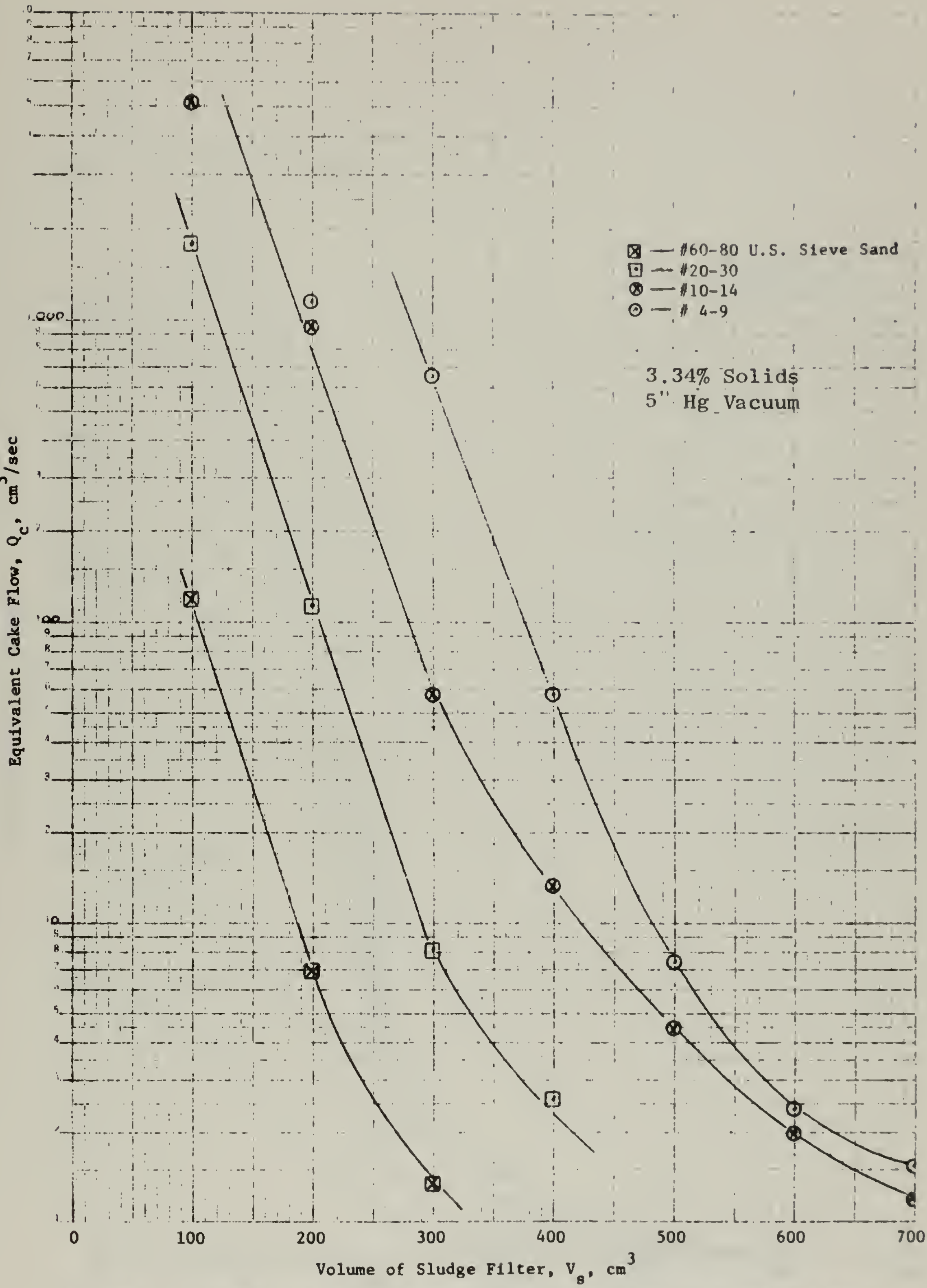

Figure 4-4. Equivalent Cake Flow as a Function of the Volume of Sludge Filtered.

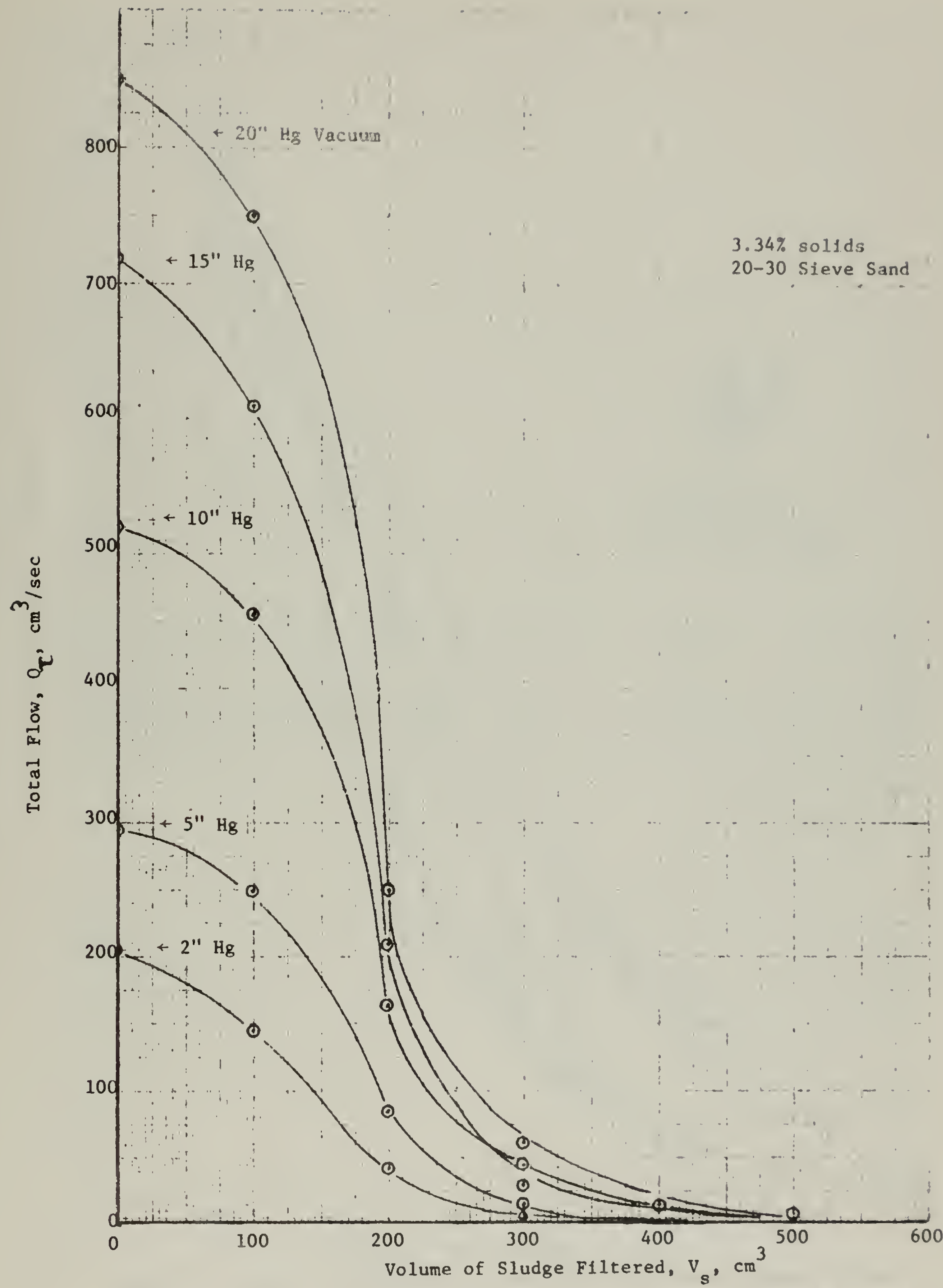

Figure 4-5. Total Flow as a Function of the Volume of Sludge Filtered.

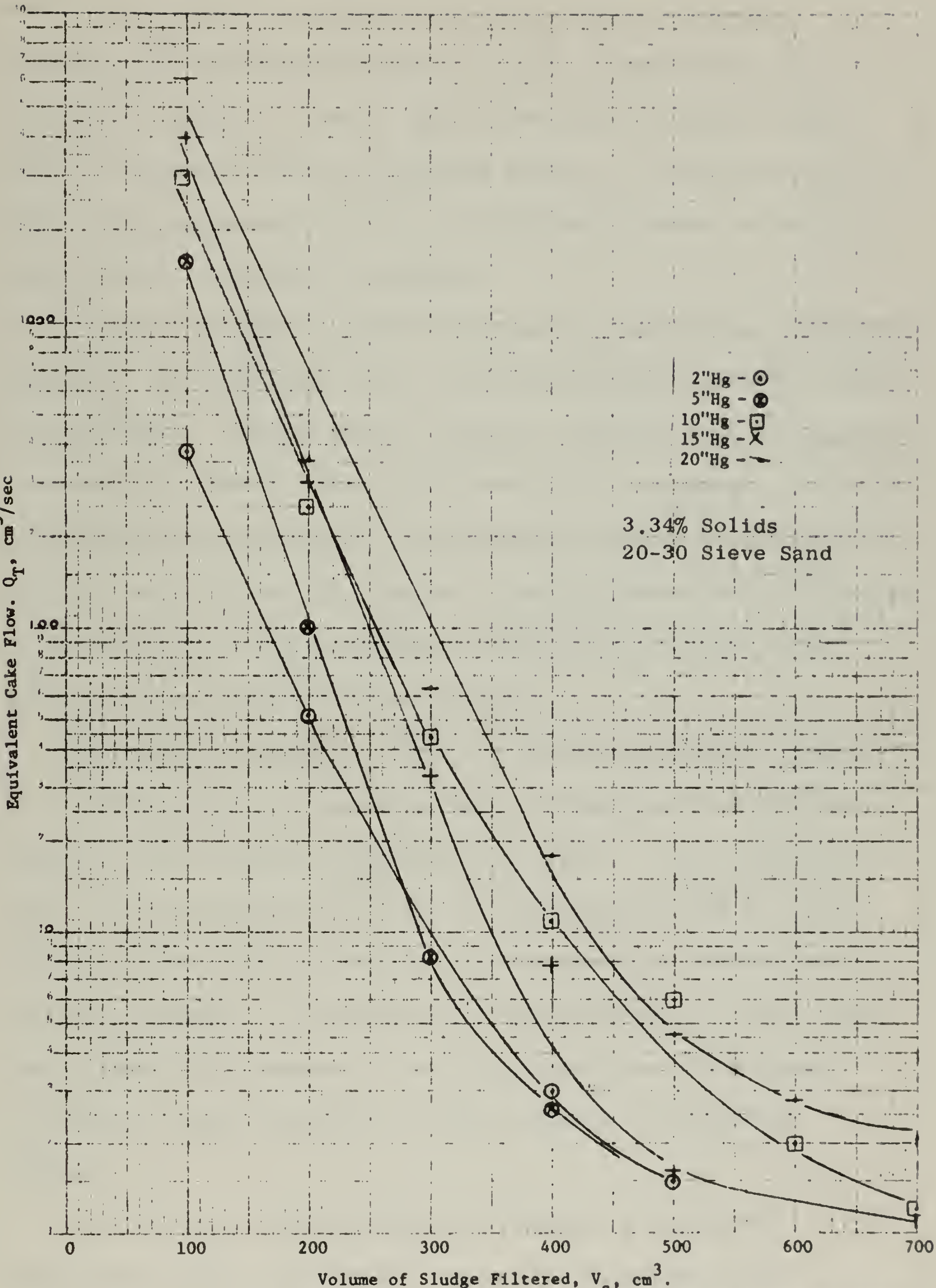

Figure 4-6. Equivalent Cake Flow as a Function of the Volume of Sludge Filtered.

can be noted that the increased flow rates produced by increasing the vacuum are quite significant for the initial incremental volumes of sludge filtered but fall rapidly at nearly the same value of V_s, 200 ml. Thus, it can be concluded that while higher initial filtration rates are produced because of the greater driving force which increases Q_a and Q_c, no significant change in the filter cake specific resistance is produced.

Figures 4-7 and 4-8 present the results obtained from filtering sludges on the 20-30 mesh medium with three different initial solids concentrations. The raw primary sludge as obtained from the Boulder treatment plant had a solids concentration of 3.34 percent. The 5.30 percent solids concentration was produced by concentrating the 3.34 percent sludge and the 1.85 percent solids concentration was obtained by the dilution of the 3.34 percent sludge with the clear liquid supernatant from the centrifuged sludge.

The curves in Figures 4-7 and 4-8 indicate an apparent increase in filtration rate for a given volume of sludge filtered with decreasing initial solids concentrations; however, the increased flow rates of the filtrate are off set by the decreased rate of solids built up on the filter medium. A greater volume of sludge must be filtered to obtain a given amount of sludge solids. In other words, the filtrate flow capacity of the filter would have to be greater to separate a given quantity of solids from their associated liquid.

Figure 4-9 is a plot of Q_T as a function of the solids deposited on the filter medium. The solids deposited were

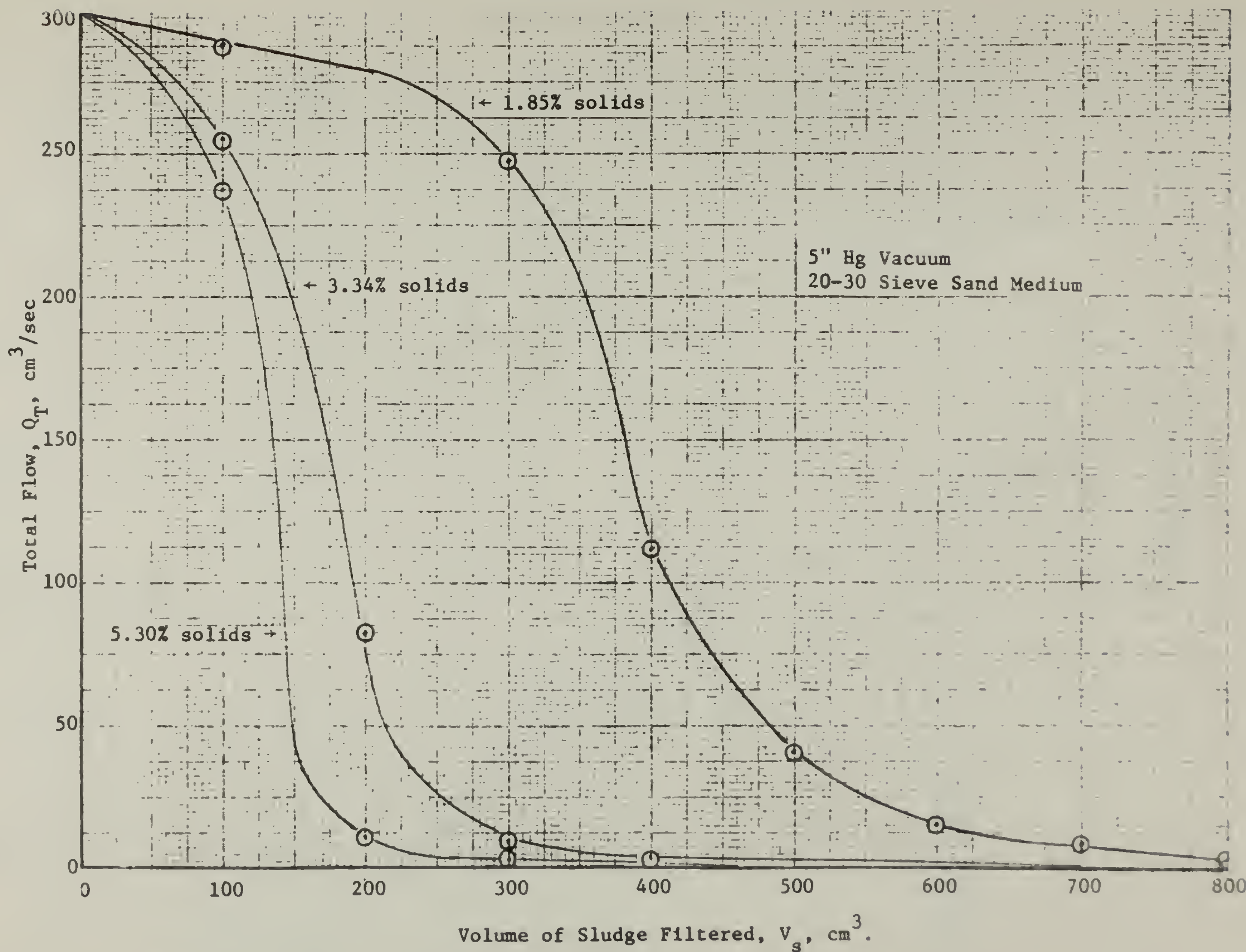

Figure 4-7. Total Flow as a Function of the Volume of Sludge Filtered.

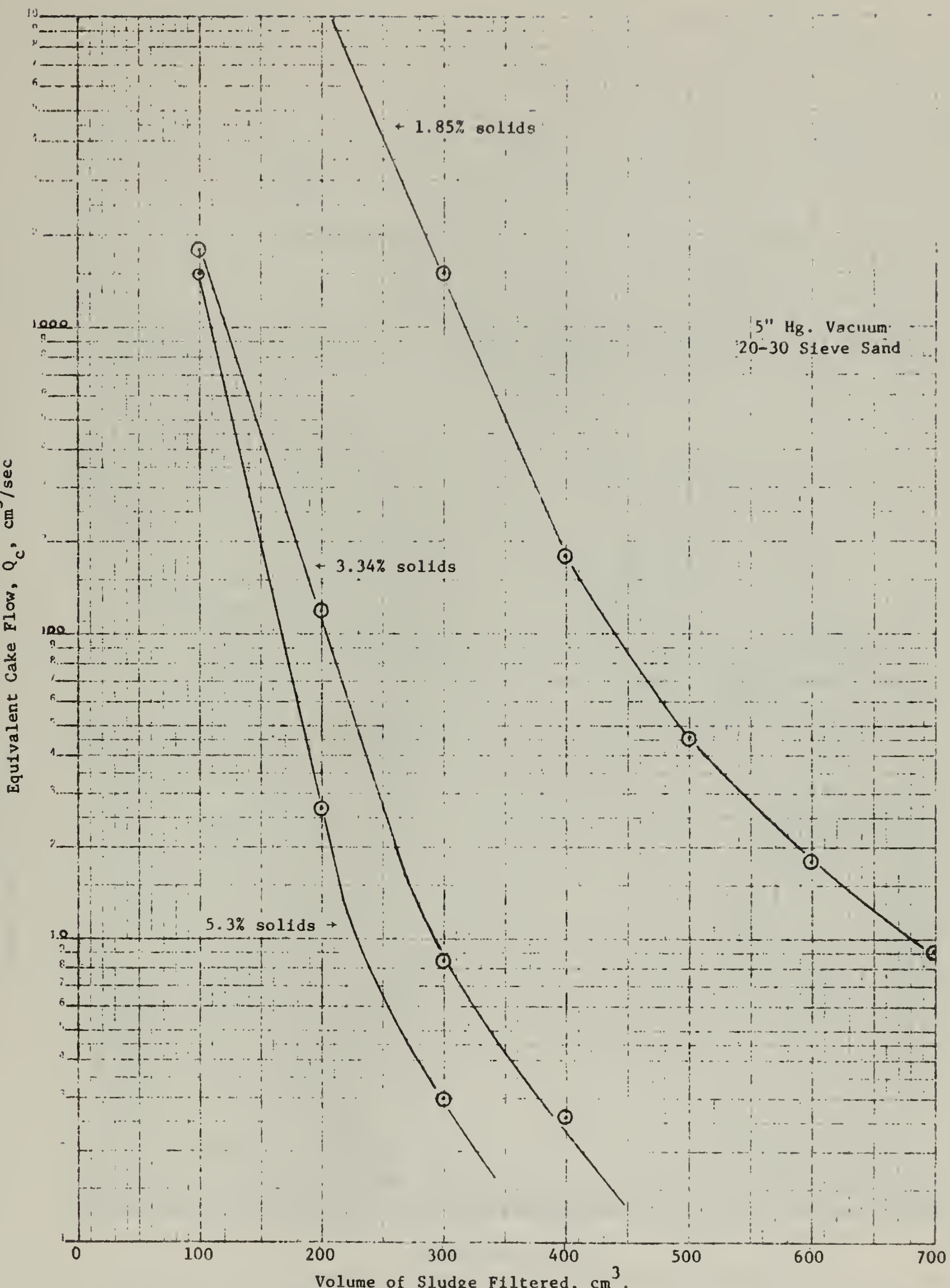

Figure 4-8. Equivalent Cake Flow as a Function of the Volume of Sludge Filtered.

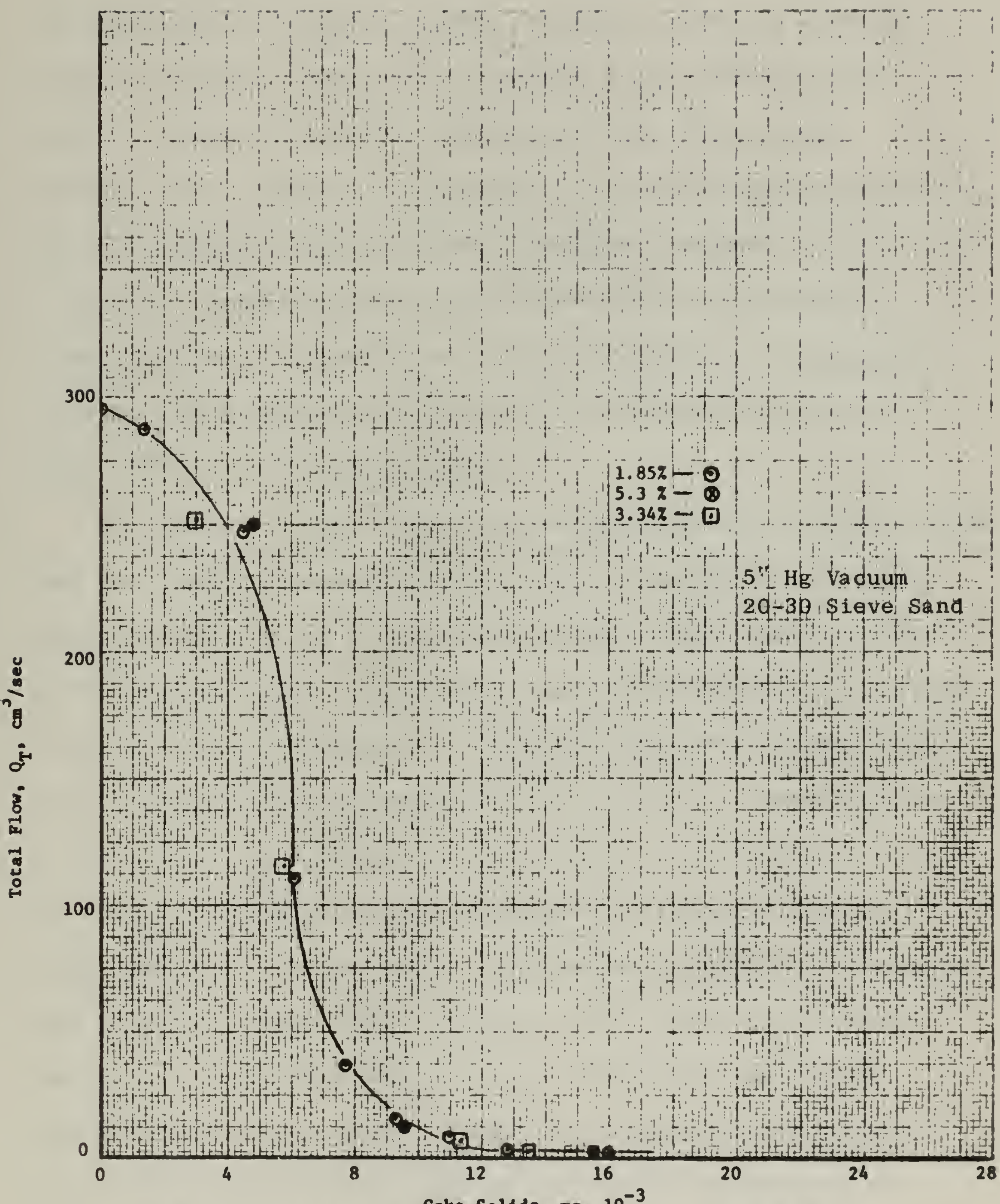

Figure 4-9. Total Flow as a Function of the Cake Solids Deposited.

computed from a solids balance between the suspended solids applied to the filter and the suspended solids in the filtrate for the three sludges used to obtain Figures 4-7 and 4-8. The fact that the three curves plotted in Figure 4-7 plot as the same curve in Figure 4-9 clearly shows that the initial solids concentration does not affect the specific resistance of the filter cake. Dilution merely decreases the quantity of solids deposited per unit volume of sludge filtered and thus, produces a thinner cake with a lower total resistance to filtrate flow for a given volume of sludge filtered. Concentration of the sludge prior to filtration actually lowers the cost of filtration by reducing the volume of liquid to be handled by the filter.

The results of this study show that much greater filtration rates than are found in present practice can be obtained for raw sewage sludge with the use of a large pore diameter filter medium. The actual increase in filtration rates that can be obtained will depend on the filter cake thickness. Considering that the minimum thickness of cake that can be removed from the filter is one-sixteenth of an inch (4), an extremely high filtration rate is possible if the filter is designed with a low apparatus resistance and if a large fraction of solids can be tolerated in the filtrate. Assuming that some fine solids can be treated by other means in the treatment plant, the problem becomes one of determining the percent solids separation desired to produce the most economical process. The filter medium pore size opening must then be determined to obtain the desired removal based on the particle size distribution

of the sludge. Chemical conditioning could be used to alter the particle size distribution. This would alter the relationship between the filter medium and filtrate flow and would require a reevaluation of the minimum cost considerations.

Review of Plant Scale Operations

During the early phases of this study the author spent some time becoming familiar with the vacuum filtration operation at the Boulder treatment plant. The unit studied was a Komline Sanderson Coilfilter. Various measurements were made while operating the filter which lead to the following conclusions:

(1) The filter as designed did not provide the operator with sufficient cycle time control; cake thickness could not be sufficiently varied to obtain optimum results.

(2) The operator could not control the form vacuum; the same vacuum source served both cake dewatering and cake formation and early cracking of the cake during dewatering produced a loss of form vacuum.

(3) The cake dewatering portion of the cycle drained only a small amount of moisture from the cake pores which lead to the conclusion that the air pulled through the cake had little effect on final cake moisture.

(4) Adequate operation of a filter requires a minimum of filtration knowledge; however, to obtain optimum results a greater understanding of the interrelationship of the process variables is required.

(5) The raw sludge at the Boulder plant can be vacuum filtered to a moisture content of approximately 80 percent and maximum yield without conditioning chemicals. This is possible because the fine solids pass through the filter medium and are returned to the plant. The addition of conditioning chemicals reduced the solids concentration in the filtrate, and the odor potential of the sludge, but it did not noticeably improve the yield or final moisture content of the filter cake.

CHAPTER V

CONCLUSIONS AND RECOMMENDATIONS

One of the major problems to be solved in the operation of waste water treatment plants is the disposal of the large quantities of sludge which accumulates in different stages of the process. The increasing population, the trend towards urbanization, the more stringent effluent standards, and modern secondary waste water treatment processes have all added to the magnitude of the sludge handling and disposal problem.

There are many alternative unit operations which can be employed in solids handling and disposal systems. Since the cost of these systems constitutes a significant portion of both the total capital and operating costs of primary and secondary treatment plants, it is quite important that the alternatives be investigated to develop the most economical solids disposal system. The alternative unit operations which can be employed in the solids handling and disposal system are presented and described in Chapter 2.

The vacuum filtration of waste water sludges is a common unit operation employed in many solids handling and disposal systems. It is used to dewater the sludges to facilitate final solids disposal. The modern vacuum filters using synthetic or metallic filter media have been found to produce much higher filtration rates than the

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cloth medium filters.

The laboratory analysis of the vacuum filtration of waste water sludge was undertaken to investigate how the filter medium pore diameter effected the filtration operation. Different sized rounded sands were used as the medium in a controlled filtration apparatus to determine filtrate flow characteristics as the filter cake formed. Various initial solids concentration and vacuum levels were used in the tests to investigate their influence on filtration through the different media. The conclusions from this investigation are given:

1. Analysis of the results of the laboratory investigation of the vacuum filtration of a raw waste water sludge with various pore size filter media shows that the theoretical hyperbolic relationship between filtrate flow and volume of sludge filtered was invalid when large pore filter media were used. The filtration rates which were produced by the different media varied as the pore diameter of the medium. The larger pore diameter of the coarse filter media allowed a greater fraction of the fine sludge particles to pass through the filter with the filtrate; thus these media produced lower filter cake specific resistance values which resulted in higher filtration rates.

2. Analysis of the results of the investigation of form vacuum and initial sludge solids concentration showed that these parameters did not effect the filter cake characteristics as did a variable filter media pore size; however, they did influence filtration rates by other means. As the form vacuum was increased,

the filtration driving force increased resulting in an increase in the rate of filtration. Concentration of the sludge prior to filtration resulted in a decrease in the amount of liquid to be handled by the filter to obtain a given quantity of solids and did not change the final cake characteristics. These results confirm previous investigations (3), (39).

3. The initial high filtration rate as predicted by the hyperbolic relationship between the rate of filtrate flow and volume of sludge filtered can be neglected, as in the past, for those filter media that retain nearly all of the suspended solids. When coarse filter media, which allow various fractions of the fine suspended solids to remain in the filtrate, are used the initial high filtration rate assumes new importance and should be investigated in the filtration analysis. In contrast to the short duration of this phenomenonfor fine cloth media the duration for synthetic or metallic media of various pore size can be extended to any length from a minimum by retaining all of the solids to a maximum by allowing all of the solids to remain in the filtrate.

4. Filter medium pore size can be used as an alternative to chemical conditioning as a means of improving vacuum filtration rates of raw waste water sludges. Since the major operational cost of vacuum filtration is the chemical conditioning cost, waste treatment plants may find it more economical to recycle a portion of the sludge solids back to the treatment plant rather than to condition the sludge prior to filtration.

5. The procedure to be followed in designing vacuum filters for

waste water sludges under the filter medium pore size control concept would be: (a) determine the optimum, economic balance between suspend solids recycle and filtration rate, (b) determine the minimum cake thickness that can be removed from the filter medium and insure adequate drum speed to maintain this cake thickness during filtration, (c) determine the particle size distribution of the sludge, and (d) determine the filter medium pore size which will give the desired filtration rate and solids recycle. It may be necessary to use chemical conditioning as a means of altering the particle size distribution of the sludge to obtain the desired results. Vacuum filtration approached in this manner is essentially a solids classification operation; those solids that can be economically separated from their associated liquid are removed while the remainder are recycled.

Future Work

The experimental results indicate certain advantages that can be gained by recycling various fractions of the syspended solids that are applied to the filter. The high filtration rates which result from filtering through coarse filter media can be attributed to a lowering of the specific resistance of the filter cake in the initial stages of formation by allowing some of the fine particles to remain in the filtıate. Future investigations into this means of improving filtration rates should be conducted, for it provides an alternative to the costly process of chemical conditioning.

It is very likely that pore size control may improve the economics of the filtration operation.

Future laboratory investigation of filter medium pore size control should be conducted to develop definitive relationship between filter medium pore diameter, sludge particle size distribution, and filtration rates. The laboratory apparatus as shown in Figure 4-1 should be modified to provide more control of filtrate flow and vacuum application. Synthetic filter media with various, known pore diameters should be used in these investigations. It may also be necessary to use a synthetic sludge of known particle size distribution to actually formulate the desired relationships. Investigations of this nature should lead to design parameters whereby the filter is designed to meet sludge characteristics in contrast to merely altering sludge characteristics by conditioning to meet filter requirements. The effect of filter media size in conjunction with chemical conditioning should be studied to understand the interrelationships of these two means of reducing cake resistance to produce optimum conditions for dewatering. This type of study should be conducted on all of the different sludge classifications produced in sewage treatment plants. From these studies a procedure should eminate that could be used to evaluate the filter design and operational parameters to be used at individual sewage treatment plants.

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APPENDIX A

SAMPLE - LABORATORY DATA AND COMPUTATIONS

30 media, 3.34% sludge, 5" Hg

V_s,ml	h,"Hg	t,sec	V_f^*,ml.,	Suspended Solids in V_f,mg/l
0	5	1.9	570	1000
100	"	1.4	355	855
200	"	3.6	300	680
300	"	26.8	215	
400	"	55.7	145	
500	"	139.9	120	
600	"	123.5	80	
700	"	90.6	55	78.3

*Total volume of liquid filtered.

$$Q_a = V_f/t = 570/1.9 = 295 \text{ cm}^3/\text{sec}$$

For V_s = 200 ml.

$$Q_T = V_f/t = 300/3.6 = 83.5 \text{ cm}^3/\text{sec}$$

$$Q_c = \frac{Q_T Q_a}{Q_a - Q_T} = \frac{(83.5)(295)}{295-83.5} = 117 \text{ cm}^3/\text{sec}$$

Solids Applied = (3.34)(100) = 334

Solids Passed = (.068)(300) = 20.4

Fraction of applied solids passed = 20.4/334 = .061

APPENDIX B

SUMMARY OF RESULTS

Initial Solids Concentration Effect

30 media, 1.85% sludge, 5" Hg, Q_a = 295 cm^3/sec

V_s,ml	Q_T,cm^3/sec	Q_c,cm^3/sec	V_f,ml	Percent of Applied Solids Passed
100	287	10,560	575	28.5
200	385	-	500	22.0
300	247	1,520	495	22.0
400	112	181	360	16.9
500	40.5	46.7	360	13.5
600	16.1	17.0	330	9.0
700	9.05	9.34	310	
800	4.30	4.35	295	
900	3.05	3.08	225	7.5
1000	2.19	2.20	195	

30 media, 5.30% sludge, 5" Hg, Q_a = 295 cm^3/sec

V_s,ml	Q_T,cm^3/sec	Q_c,cm^3/sec	V_f,ml	Percent of Applied Solids Passed
100	250.0	1,640	400	11.0
200	12.2	12.7	350	9.5
300	3.0	3.03	335	3.5
400	0.596	0.596	190	1.0
500	0.575	0.575	95	0.5

30 media, 3.34% sludge, 5" Hg, Q_a = 295 cm^3/sec

V_s,ml	Q_T,cm^3/sec	Q_c,cm^3/sec	V_f,ml	Percent of Applied Solids Passed
100	254.0	1,830	355	13.5
200	83.5	117	300	10.3
300	8.05	8.26	215	6.3
400	2.60	2.62	145	
500	0.86	0.86	120	
600	0.65	0.65	80	
700	0.61	0.61	55	0.4

V_s, ml	Q_T, cm^3/sec	Q_c, cm^3/sec	V_f, ml	Percent of Applied Solids Passed

Vacuum Level Effect

30 media, 3.34% sludge, 2" Hg, Q_a = 228 cm^3/sec

100	145	398	465	14.1
200	42.7	52.6	355	10.7
300	8.15	8.45	370	6.8
400	2.88	2.92	195	
500	1.53	1.54	140	
600	0.87	0.870	140	20.4

30 media, 3.34% sludge, 10" Hg, Q_a = 520 cm^3/sec

100	456	3,710	640	10.6
200	166	244	550	9.5
300	40.4	43.6	480	8.8
400	10.8	11.0	270	
500	5.97	6.05	280	
600	1.96	1.96	170	
700	1.14	1.14	125	

30 media, 3.34% sludge, 15" Hg, Q_a = 720 cm^3/sec

100	609	3,940	730	16.2
200	218	312	545	9.8
300	29.1	30.3	320	4.3
400	7.34	7.41	365	
500	6.05	6.11	215	
600	1.58	1.58	175	0.8
700	1.14	1.14	85	
800	0.40	0.40	55	

30 media, 3.34% sludge, 20" Hg, Q_a = 850

100	750	6,360	750	14.2
200	252	358	580	10.1
300	59.6	64.2	340	3.6
400	17.5	17.9	515	
500	4.6	4.62	310	
600	2.86	2.86	165	
700	2.10	2.10	140	0.8
800	1.10	1.1	140	

V_s, ml	Q_T, cm^3/sec	Q_c, cm^3/sec	V_f, ml	Percent of Applied Solids Passed
		Filter Media Effect		
		4-9 media, 3.34% sludge, 5" Hg, Q_a = 310 cm^3/sec		
100	305	18,900	305	14.5
200	245	1,170	220	14.5
300	212	671	360	11.3
400	50	59.6	290	
500	7.30	7.46	180	
600	2.40	2.41	65	
700	1.57	1.57	60	
800	.95	.95	115	.4
		10-14 media, 3.34% sludge, 5" Hg, Q_a = 310 cm^3/sec		
100	292	5,020	380	10
200	235	972	235	6.4
300	50	59.6	345	6.1
400	12.8	13.4	305	
500	4.5	4.56	190	
600	1.94	1.95	125	
700	1.12	1.12	70	
800	.82	.82	100	.635
		60-80 media, 3.34% sludge, 5" Hg, Q_a = 193 cm^3/sec		
100	108	102	335	2.7
200	28	7.16	155	2.6
300	1.83	1.36	110	0.8
400	0.84	.48	80	
500	0.40	.12	60	0.3